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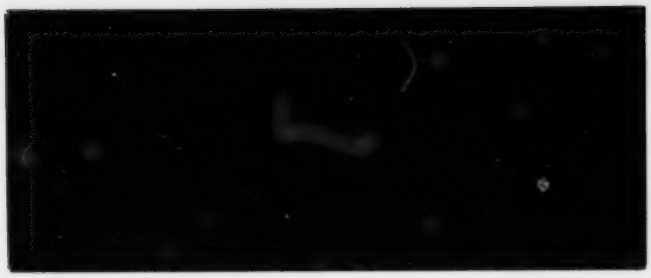

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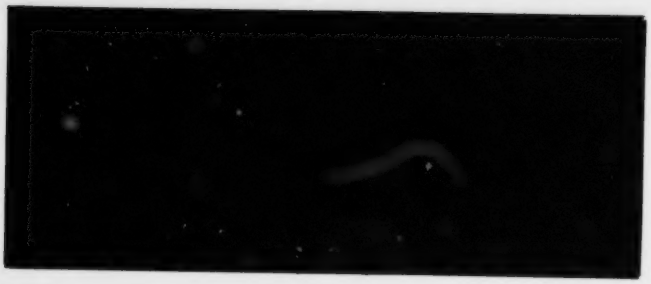

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INVESTIGATION OF SULPHUR CAPTURE DURING CIRCULATING FLUIDIZED BED COMBUSTION OF EASTERN CANADIAN COALS (Phase II)

FINAL REPORT VOLUME I

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SUMMARY

This is the Final Report on Phase II of the Investigation of Sulphur Capture During Circulating Fluidized Bed Combustion of Eastern Canadian Coals.

Minto coal was tested with Havelock, Elmtree and Albert county oil shale. Prince coal was tested with Havelock, Glencoe and Windsor limestones.

In this study all the significant parameters affecting sulphur capture have been examined. The parameters tested were limestone type and particle size, Ca/S ratio, bed temperature, operating velocity, secondary cyclone recycle, and solid circulation rate. The effect of each of these parameters on sulphur capture has been identified. Pressure drop measurements were carried out and the effect of the J-valve performance on the solids loading in the combustor was examined.

Burning Minto coal with Havelock limestone at a Ca/S ratio of 2:1, sulphur emissions were reduced by 87% This was possible only with recycling of the dust from the secondary cyclone back into the combustor. At the same Ca/S ratio and without dust recycling, the highest attainable sulphur capture was only 76%. For Prince coal and Windsor limestone higher sulphur capture was possible, using the same Ca/S ratio and without dust recycling, a sulphur capture of 93% was achieved.

The study met all of its objectives. Correlating of these results with those from N.B. Power demonstration unit at Chatham will be made when these become available. A number of important conclusions have been drawn as well as several recommendations made for future work. These include, identification of the most effective limestone particle size range, highlighting the importance of dust recycle and ranking of the sorbents tested. One of the important future tasks would be the evaluation of different options for improving sulphur capture and sorbent utilization. This may include increasing the reactivity of natural sorbents and fly ash. Correlating the performance of RPC pilot scale unit with the Chatham demonstration unit is perhaps the most important next task.

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1.0 INTRODUCTION

The Research and Productivity Council was contracted by the Federal and Provincial governments to carry out a comprehensive study of sulphur capture during the circulating fluidized bed combustion (CFBC) of Eastern Canadian coals. These are high sulphur coals and new technology that will limit sulphur oxides (SO_x) emissions is required to comply with the air quality guidelines. The project was funded under the Industrial Innovation and Technology Development Subsidiary Agreement of the province of New Brunswick and the federal department of Energy, Mines and Resources. The federal funding was from the 50/50 cost shared Energy Conversion Program and from the Program of the Federal Panel on Energy Research and Development (PERD).

In the first phase of the work carried out in 1987 the RPC pilot plant CFBC was modified so that the test results were applicable to the operation of the 20 MW demonstration unit at Chatham, NB, operated by N.B.Electric Power Commission. The results of that work on equipment modification and equipment testing are given in the RPC report MDP/87/04, "Investigation of Sulphur Capture During Circulating Fluidized Bed Combustion of Eastern Canadian Coals - Phase I" by R. Robson and P. Salib.

The results of the second part of the work is the subject of this report. It describes the combustion tests on New Brunswick coal (Minto) with Elmtree and Havelock limestones and the Albert county oil shale (Tasks I-V) and on Nova Scotia coal (Prince) and limestones from Glencoe and Windsor (Task V). The five tasks were as follows:

Task I — To determine the effect of secondary cyclone recycle and limestone particle size on sulphur capture.

Task II — To determine the effect of limestone type (Havelock and Elmtree), particle size, Ca/S ratio and gas velocity on sulphur capture. The effect of secondary cyclone recycle was also re-examined.

Task III To determine the effect of circulating rate and bed temperatures on sulphur capture by CFBC. Secondary cyclone recycle was once more tested. Step changes were also made in limestone feeding to obtain information on the dynamic behaviour of the systems.

Task IV To determine the effectiveness of oil shale as a sulphur capture agent. The influence of shale particle size, shale/coal ratio and Ca/S ratio were determined.

Task V To determine the sulphur capture attainable during the combustion of Prince Coal with Glencoe and Windsor limestones. The variables tested were Ca/S ratio, limestone type and particle size.

The tasks outlined above have been completed successfully in fulfilment of the statement of work given in Appendix A. It is worth noting, however, that the Prince coal work which is presented in this report as Task V is, according to the statement of work, part of Task II. This was due to a delay in receiving Nova Scotia feedstock, hence the test work was performed at a later date and the chemical and data analysis were carried out and compiled as a separate task.

Volume I of this report describes the work carried out including test and measurement procedures, results and discussion and conclusion and recommendation for future work. Volume II contains all the raw data chemical analysis and mass balances.

2.0 TEST FACILITY

In Phase I, the combustor was successfully modified and tested. The reason for the modification was to reduce bed inventory in order to make it compatible with commercial designs. This modification also reduces the time required for steady state operation. More details on bed modification and the associated experimental work can be found in (1).

The modified facility is shown schematically in Figure 1. The main components of the system are the combustor, a primary cyclone and a standpipe for solid circulation, a cooling section, secondary cyclones and a baghouse.

The Combustor

The cylindrical combustor has a total height of 5.8 m (19') and inside and outside diameters of 12.7 cm (5") and 50.8 cm (20") respectively. The wall has an inner layer of high temperature refractory, a middle layer of castable insulating refractory, and a carbon steel shell. A 0.5 MBtu propane burner is used for start-up purposes.

Primary air is fed into the bottom of the combustor via a distributor plate; a perforated plate with 10% open area located 66 cm (26") from the base of the combustor column. Secondary air is fed into the combustor via four diagonal ports at 205.7 cm (81") above the distributor plate.

Stand Pipe

The standpipe is used for storage and recirculation of solids received from the primary cyclone. It is 0.9 m high and has inside and outside diameters of 22.8 cm (9") and 62 cm (24") respectively. The standpipe is refractory lined and is equipped with discharge and sampling ports.

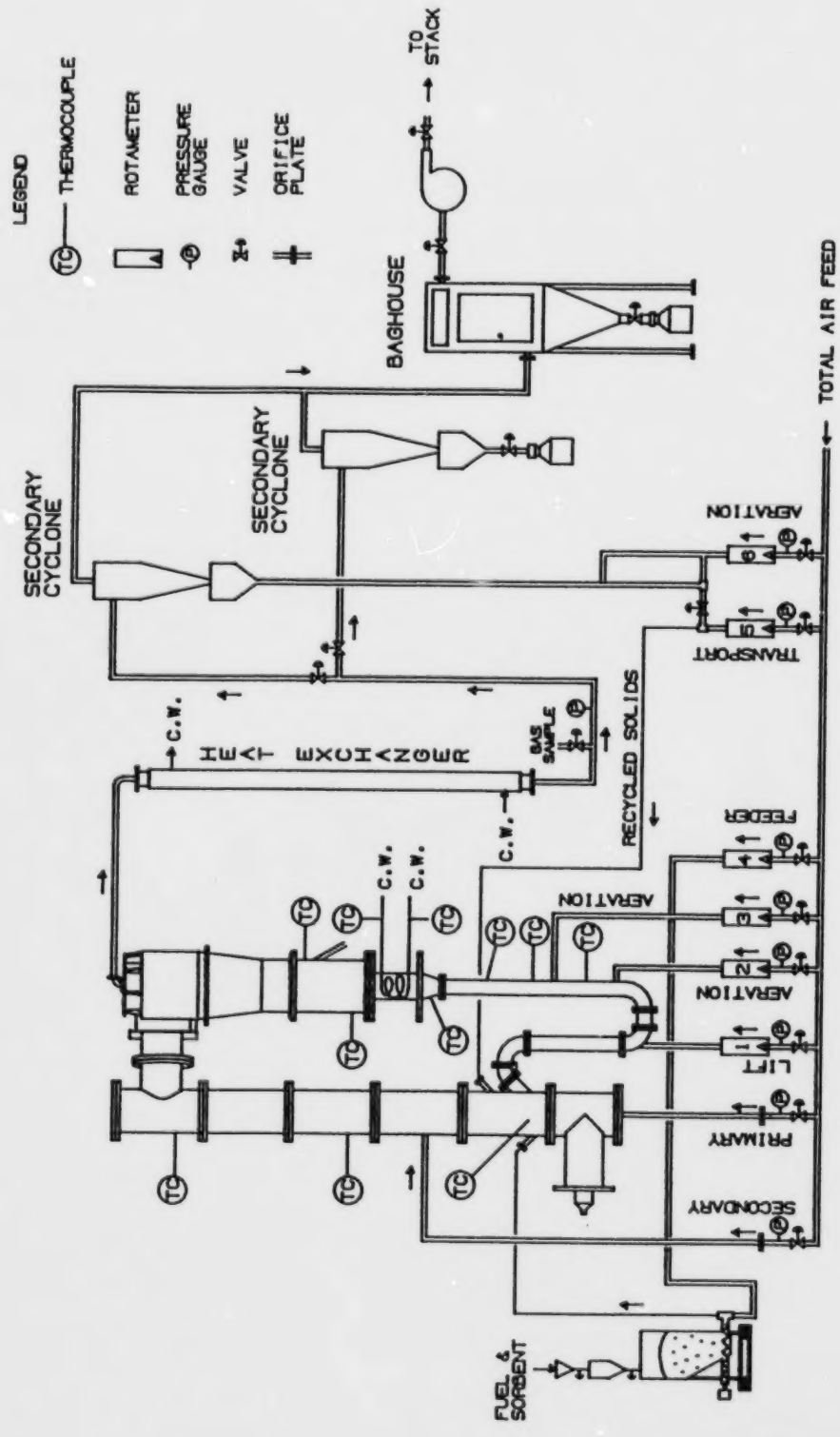

FIGURE 1

SCHEMATIC OF RPC CIRCULATING FLUIDIZED BED COMBUSTOR SHOWING TEMPERATURE, ABSOLUTE PRESSURE AND FLOW SENSORS

A cooling section is installed at the bottom of the standpipe. This is used to extract excess heat generated in the combustor so that both bed temperature and air/fuel ratio can be set and maintained at the appropriate levels.

Cyclones

The primary cyclone is 30 cm (12") in diameter, and is refractory lined in a similar manner to the combustor and the standpipe. The two identical secondary cyclones are made of stainless steel and are 25 cm (10") in diameter. The solids collected by the secondary cyclone can be recycled pneumatically to the combustor via a port, 48.3 cm (19") above the distributor plate.

The J-Valve

Circulation of solids from the standpipe to the combustor is achieved by a non-mechanical J-valve. The vertical section is made of 3" Schedule 40 stainless steel pipe. It is aerated at two points approximately 0.21 and 1.06 m from the base of the J-valve. The curved section is made of 2" schedule 40 stainless steel pipe. Lift air is supplied at the base in order to transport the solids into the combustor entering at a location 0.37 m above the distributor plate.

Gas Cooling System

Off-gases from the primary cyclone are fed into a water cooled one pass heat exchanger. It is made of two concentric pipes of a nominal diameter of 15 cm (6") (Schedule 40) and 20 cm (8") (Schedule 40), respectively. The height of the heat exchanger is 5.2 m (17').

The Baghouse

The flue gases from the cooling section are fed through the secondary cyclone into the baghouse. The latter is a continuous duty self cleaning dust collector which consists of 16 high temperature bags capable of handling 200 SCFM.

Solid Feeding

Dry solids are fed pneumatically into the combustor 44.5 cm (17.5") above the distributor plate. The feed rate is regulated using a screw feeder. The feeder assembly is mounted on a weighing scale to provide an accurate record of the feed rate.

3.0 START-UP, TEST AND MEASUREMENT PROCEDURES

The system is preheated using a 0.5 MBtu/hr propane burner, which is normally fired at half its rating. Once the combustor temperature has reached 200-300°C, the bed material is introduced into the system. For this particular test program ash from previous tests was used as the starting bed material. Coal feed was initiated and the propane burner switched off at a bed temperature of approximately 600°C. At a bed temperature of approximately 850°C, the cooling section at the bottom of the standpipe was activated and the operating conditions were set for a particular test run.

Tests were carried out sequentially, with a transition period between tests of 2 to 6 hours. Gas composition (particularly SO_2 concentration) and bed temperature were the two main indicators used to signal the end of the transition period. Test duration was nominally 24 hours. For each test the experimental measurements and analysis consisted of the following:

Solids Flow Rates

(a) Feed, Standpipe Overflow, Baghouse Dust

These measurements were taken every hour.

(b) Secondary Cyclone Recycle

This measurement was taken every four hours in order to minimize the disruption to the circuit. To take this measurement, the stand-by secondary cyclone (see Figure 2) was brought on-line and the underflow was collected for a given period of time (typically 5 min).

(c) Solids Recycle Rate Through Standpipe

The recycle rate was calculated every hour by carrying out a simple heat balance calculation as illustrated in Figure 3.

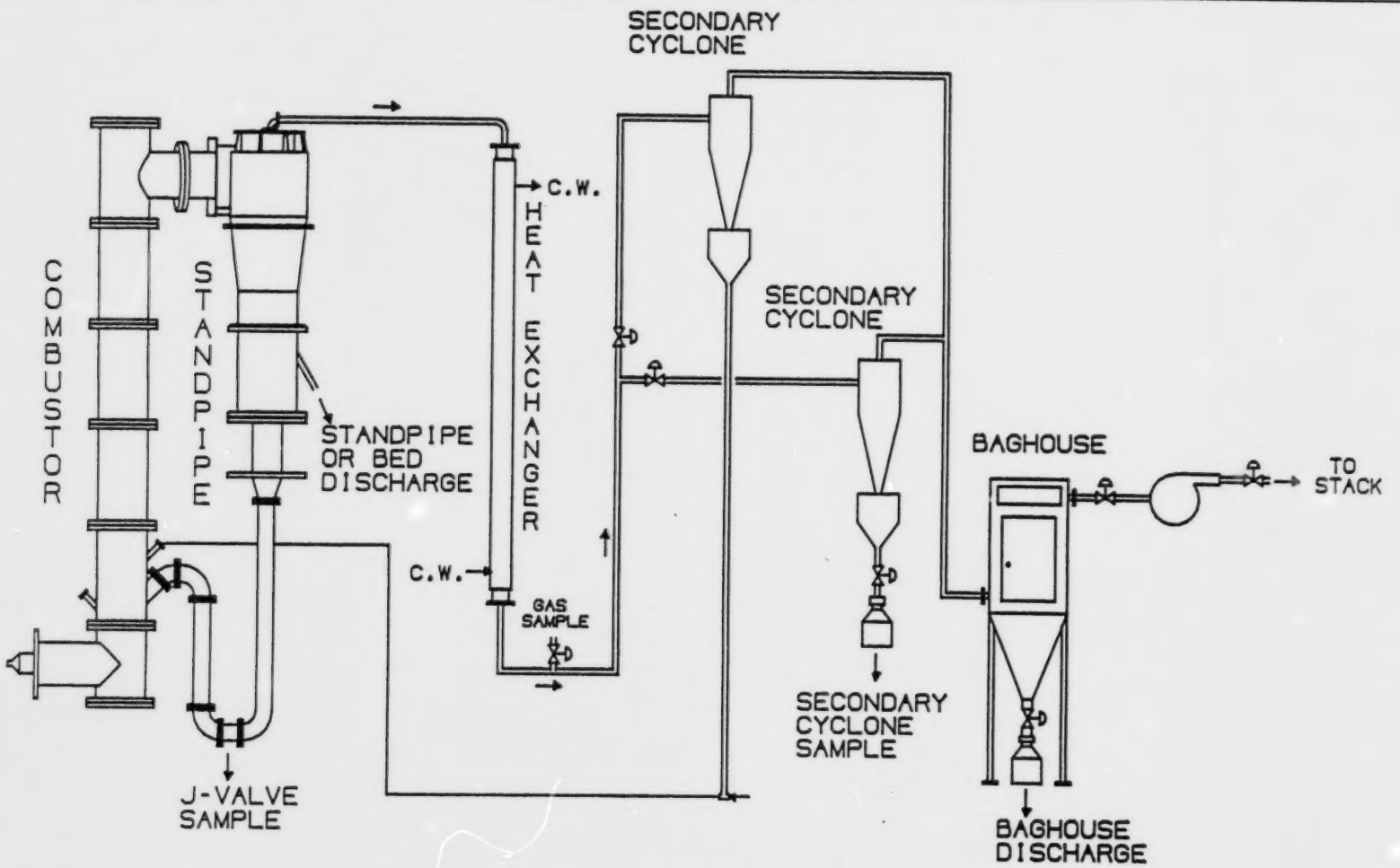

FIGURE 2

SCHEMATIC OF RPC CIRCULATING FLUIDIZED BED COMBUSTOR SHOWING SOLID AND GAS SAMPLING LOCATIONS

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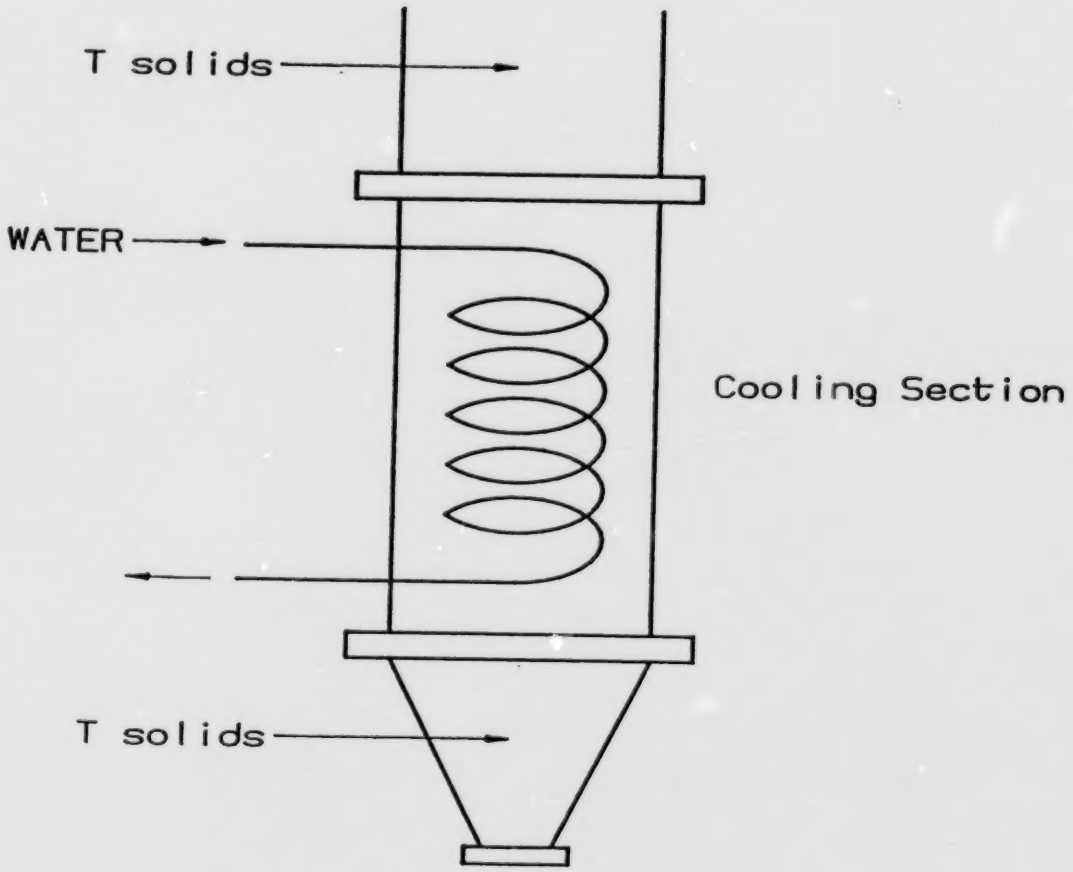


$$\text{mass flow of solids} = \frac{Cp_{water} \times \Delta T_{water} \times \text{mass flow water}}{\Delta T_{solids} \times Cp_{solids}}$$

FIGURE 3

DETERMINATION OF THE SOLIDS RECIRCULATION RATE

Air Flow Rate

Primary air and secondary air flows to the combustor were recorded every hour. Also recorded at hourly intervals were lift and aeration air to the J-valve, feeder air and transport and aeration air for the secondary cyclone recycle.

Temperatures

Temperatures were recorded hourly at several locations in the combustor and the standpipe.

Pressures

Differential pressures were recorded at nine locations in the combustor and four locations in the standpipe as shown in Figure 4.

Gas Sampling and Analysis

Continuous on-line monitoring for CO, CO_2, O_2, SO_2 and NO_x took place with the averages recorded hourly. The sampling location was downstream of the heat exchanger as shown in Figure 2.

Solids Sampling

Four solids flows were sampled at four-hourly intervals. These were J-valve, standpipe overflow, secondary cyclone underflow and baghouse dust. When the secondary cyclone is removed from the circuit, the number of samples were reduced to three. In addition, samples of coal and limestone were taken at the start of each test period.

Solids Sample Analysis

A run was of a nominal twenty-four hour duration. Three samples were submitted for analysis at 8 hour intervals from the initiation of a test. All three samples were analyzed for total carbon, total sulphur, calcium and ash. In addition to these analyses, the third sample was analyzed for sulphate sulphur, inorganic carbon and free lime.

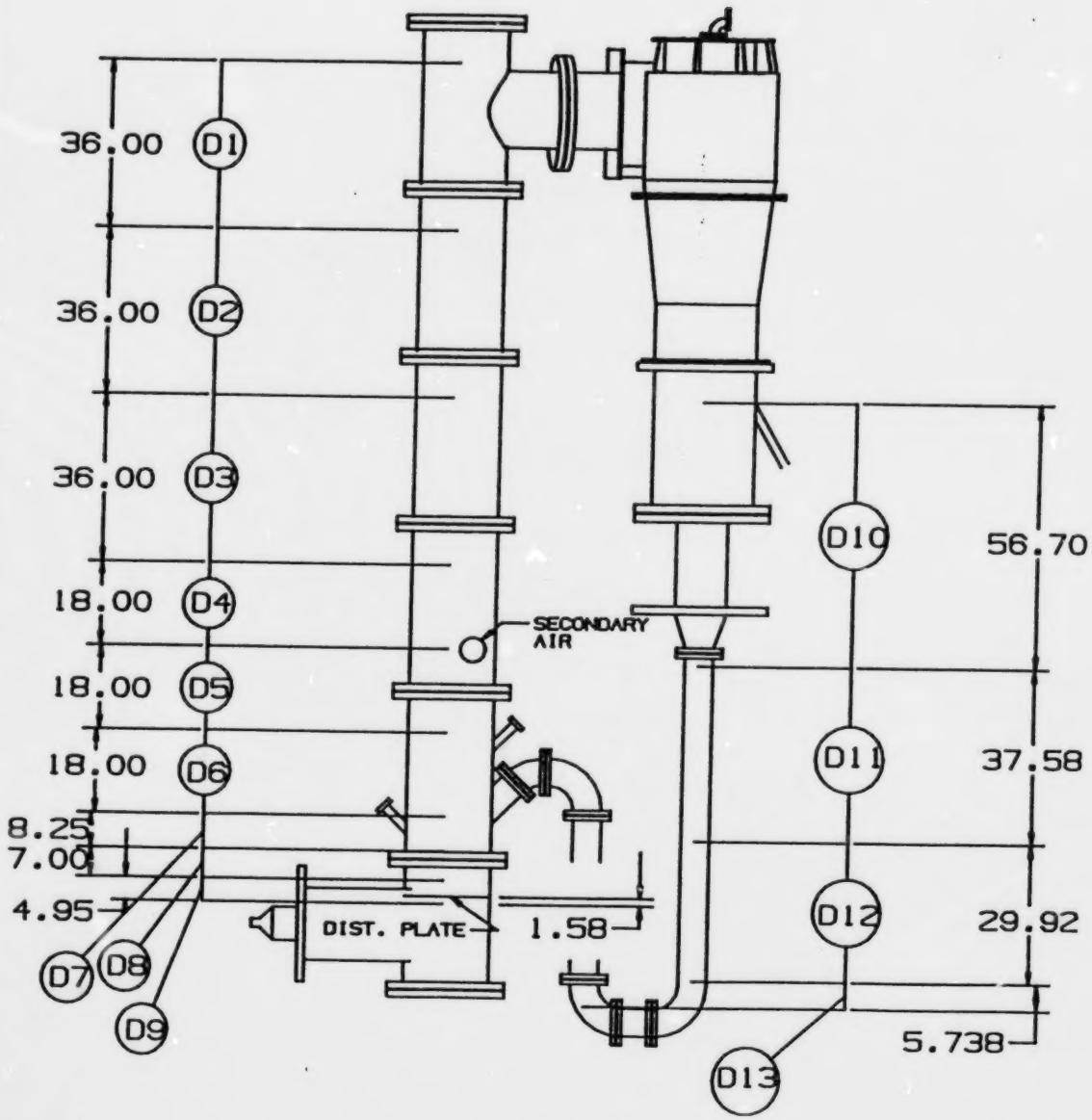


FIGURE 4

SCHEMATIC OF RPC CIRCULATING FLUIDIZED BED COMBUSTOR SHOWING THE LOCATIONS OF DIFFERENTIAL PRESSURE GAUGES

4.0 RESULTS AND DISCUSSION

The raw data, chemical analysis and mass balances for each experimental run are given in Appendix B. Details of data format and calculation procedures are also given in the Appendix. For completion, however, a set of results and a brief outline is given below.

A full set of raw data and calculated results for a single experimental run (Test II-1) is given in the following 10 pages. The information given in each page is as follows:

PAGE	
13	Title page gives task and experiment number, start and finish time, limestone type and size, etc.
14	Average values and other statistical information on the recorded test data.
15-17	Hourly recorded raw data.
18	Composition and flowrate of coal and limestone and flue gas analysis.
19-20	Composition and flowrate of the solid output streams (cyclone, j-valve, baghouse, standpipe and weepage).
21	Mass balances (overall, carbon, sulphur, calcium and ash).
22	Graphical presentation of solids feedrate, bed temperature and flue gas composition (CO_2, O_2 and SO_2).

The results were then compiled into a number of tabulations. The most comprehensive of which is Table 1 which shows all the significant operating , measured and calculated parameters for all of the experimental runs.

As can be seen from Tables 1, the majority of the experimental runs were carried out using Minto Coal, Havelock and Elmtree limestones (Tasks I to III). In the first three tasks the operating parameters affecting sulphur capture were examined. The knowledge gained was then utilized in designing the test matrices for Tasks IV and V, oil shale work and the Nova Scotia feedstock tests, respectively.

SULPHUR CAPTURE STUDY

File Name: So2TK1T2

Task No.: 2 Test No.: 1

Start: 18:00-08/12/87

Finish: 14:00-09/12/87

Limestone Type: Havelock

Limestone Size: Coarse-unscreened

Ca/S Ratio: 1.5:1

Secondary Cyclone: ON

		Avg	Std	Count	Min	Max
Solids Feedrate	kg/hr	10.94	0.97	20.00	9.00	12.20
Secondary Cyclone Sample	kg	0.06	0.13	20.00	0.00	0.35
Secondary Cyclone Discharge	kg/hr	3.75	0.26	4.00	3.60	4.20
J-Valve Sample	kg	0.15	0.30	20.00	0.00	0.90
Baghouse Discharge	kg/hr	2.71	0.61	20.00	1.10	3.70
Standpipe Discharge	kg/hr	0.54	0.78	20.00	0.10	3.70
Heat Exchanger Inlet Temp.	C	828.79	15.73	19.00	795.00	859.00
Heat Exchanger Outlet Temp.	C	776.32	18.99	19.00	741.00	808.00
Recycle Rate	kg/hr	1023.00	312.19	17.00	601.00	1595.00
Primary Air Differential	in. H20	29.45	1.71	20.00	27.00	33.00
Primary Air Pressure	psig	3.67	0.29	20.00	3.20	4.40
Primary Air Feedrate	kg/hr	59.93	1.83	20.00	57.15	63.70
Secondary Air Differential	in. H20	6.30	0.48	20.00	6.00	7.50
Secondary Air Pressure	psig	0.97	0.32	20.00	0.50	1.80
Secondary Air Feedrate	kg/hr	10.35	0.45	20.00	9.95	11.59
Total Flow Differential	in. H20	24.68	0.90	20.00	23.00	26.00
Total Flow Pressure	psig	16.03	0.66	20.00	15.50	17.50
Total Flow Feedrate	kg/hr	105.72	2.88	20.00	101.20	111.11
Rotameter 1 % Flow	%	27.30	2.45	20.00	25.00	30.00
Rotameter 1 Pressure	psig	2.69	0.77	20.00	1.50	4.00
Rotameter 1 Flow	kg/hr	3.97	0.39	20.00	3.51	4.53
Rotameter 2 % Flow	%	27.10	2.65	20.00	25.00	35.00
Rotameter 2 Pressure	psig	3.47	1.31	20.00	1.50	6.50
Rotameter 2 Flow	kg/hr	4.02	0.39	20.00	3.51	5.14
Rotameter 3 % Flow	%	20.95	2.60	20.00	15.00	27.00
Rotameter 3 Pressure	psig	3.16	1.01	20.00	1.75	6.50
Rotameter 3 Flow	kg/hr	3.09	0.38	20.00	2.20	3.96
Rotameter 4 % Flow	%	65.55	1.50	20.00	63.00	68.00
Rotameter 4 Pressure	psig	3.06	0.35	20.00	2.50	3.80
Rotameter 4 Flow	kg/hr	9.64	0.27	20.00	9.12	10.20
Rotameter 5 % Flow	%	21.13	2.02	20.00	17.50	25.00
Rotameter 5 Pressure	psig	2.02	0.41	20.00	1.50	3.00
Rotameter 5 Flow	kg/hr	3.01	0.29	20.00	2.50	3.60
Rotameter 6 % Flow	%	11.65	0.85	20.00	10.00	13.00
Rotameter 6 Pressure	psig	5.99	0.17	20.00	5.50	6.50
Rotameter 6 Flow	kg/hr	1.85	0.14	20.00	1.59	2.06
Sum Rotameters 1 to 6	kg/hr	25.58	0.83	20.00	24.12	26.98
Rotameters Total Flow Diff.	in. H20	8.00	0.27	20.00	7.50	8.50
Rotameters Total Flow Pressure	psig	15.48	0.68	20.00	15.00	17.00
Rotameters Total Flow	kg/hr	25.07	0.59	20.00	24.09	26.49
Purge % Flow	%	18.79	2.55	19.00	16.00	24.00
Purge Pressure	psig	14.53	0.17	19.00	14.00	14.80
Purge Flow	kg/hr	6.32	0.85	19.00	5.39	8.07
Sum Total Flow	kg/hr	101.36	2.79	20.00	93.08	106.85
Combustor Bottom Temp.	C	828.80	16.58	20.00	799.00	868.00
Combustor Middle Temp.	C	850.55	17.76	20.00	817.00	892.00
Combustor Top Temp.	C	851.95	17.89	20.00	818.00	890.00
Standpipe Top Temp.	C	829.65	16.16	20.00	795.00	861.00
Standpipe Middle Temp.	C	743.45	24.20	20.00	684.00	781.00
Standpipe Bottom Temp.	C	750.60	19.28	20.00	717.00	789.00
CO2 Concentration	%	14.55	0.94	20.00	13.10	16.20
O2 Concentration	%	4.39	0.63	20.00	3.00	5.50
CO Concentration	ppm	150.70	38.36	20.00	93.00	234.00
SO2 Concentration	ppm	1805.56	303.17	18.00	1250.00	2300.00
NOx Concentration	ppm	227.55	19.14	20.00	178.00	254.00

		18:00	19:00	20:00	21:00	22:00	23:00	24:00	1:00
Solids Feedrate	kg/hr	12.00	12.20	11.60	11.40	11.80	12.20	11.80	9.80
Secondary Cyclone Sample	kg	0.00	0.00	0.00	0.00	0.30	0.00	0.00	0.00
Secondary Cyclone Discharge	kg/hr					3.60			
J-Valve Sample	kg	0.00	0.00	0.00	0.00	0.70	0.00	0.00	0.00
Baghouse Discharge	kg/hr	1.10	1.90	2.60	2.80	2.80	3.30	2.40	2.10
Standpipe Discharge	kg/hr	0.10	3.70	0.20	0.10	0.20	0.10	0.40	0.60
Heat Exchanger Inlet Temp.	C		823.00	821.00	816.00	818.00	817.00	813.00	804.00
Heat Exchanger Outlet Temp.	C		799.00	764.00	755.00	759.00	753.00	743.00	741.00
Recycle Rate	kg/hr			847.00	713.00	841.00	690.00	626.00	601.00
Primary Air Differential	in. H20	27.00	28.00	28.00	27.00	27.00	27.00	28.50	28.50
Primary Air Pressure	psig	3.50	3.25	3.50	3.50	4.00	4.00	3.20	3.50
Primary Air Feedrate	kg/hr	57.15	57.80	58.20	57.15	57.93	57.93	58.23	58.71
Secondary Air Differential	in. H20	6.00	6.00	6.00	6.00	6.00	6.00	6.00	6.00
Secondary Air Pressure	psig	0.75	0.50	1.00	1.00	1.50	1.50	0.60	0.75
Secondary Air Feedrate	kg/hr	10.03	9.95	10.11	10.11	10.27	10.27	9.98	10.03
Total Flow Differential	in. H20	23.50	24.00	24.00	24.00	23.00	24.00	23.50	24.00
Total Flow Pressure	psig	15.50	15.50	15.50	15.50	15.50	15.50	15.50	15.50
Total Flow Feedrate	kg/hr	102.30	103.38	103.38	103.38	101.20	103.38	102.30	103.38
Rotameter 1 % Flow	%	30.00	30.00	30.00	30.00	30.00	30.00	30.00	30.00
Rotameter 1 Pressure	psig	4.00	1.75	2.50	2.00	2.50	2.50	3.00	4.00
Rotameter 1 Flow	kg/hr	4.53	4.25	4.34	4.28	4.34	4.34	4.40	4.53
Rotameter 2 % Flow	%	27.50	25.00	27.50	27.50	27.50	30.00	35.00	30.00
Rotameter 2 Pressure	psig	2.50	3.00	1.75	5.50	2.75	3.00	3.00	2.75
Rotameter 2 Flow	kg/hr	3.98	3.67	3.89	4.31	4.01	4.40	5.14	4.37
Rotameter 3 % Flow	%	17.50	15.00	20.00	20.00	20.00	20.00	20.00	20.00
Rotameter 3 Pressure	psig	2.50	3.00	1.75	2.00	3.00	3.50	3.50	4.00
Rotameter 3 Flow	kg/hr	2.53	2.20	2.83	2.85	2.94	2.98	2.98	3.02
Rotameter 4 % Flow	%	65.00	67.00	67.00	65.00	67.00	67.00	65.00	68.00
Rotameter 4 Pressure	psig	3.00	2.50	2.70	3.50	3.25	3.00	3.20	3.80
Rotameter 4 Flow	kg/hr	9.54	9.69	9.75	9.67	9.90	9.83	9.59	10.20
Rotameter 5 % Flow	%	17.50	20.00	20.00	20.00	20.00	20.00	20.00	20.00
Rotameter 5 Pressure	psig	2.00	1.50	1.75	2.00	2.00	2.00	3.00	3.00
Rotameter 5 Flow	kg/hr	2.50	2.81	2.83	2.85	2.85	2.85	2.94	2.94
Rotameter 6 % Flow	%	10.00	10.00	10.00	10.00	13.00	12.00	12.00	12.00
Rotameter 6 Pressure	psig	6.00	6.00	6.00	6.00	6.00	6.00	5.75	6.50
Rotameter 6 Flow	kg/hr	1.59	1.59	1.59	1.59	2.06	1.90	1.89	1.93
Sum Rotameters 1 to 6	kg/hr	24.66	24.21	25.23	25.56	26.10	26.31	26.94	26.98
Rotameters Total Flow Diff.	in. H20	7.50	7.50	8.00	8.00	8.00	8.00	8.50	8.00
Rotameters Total Flow Pressure	psig	15.00	15.00	15.00	15.00	15.00	15.00	15.00	15.00
Rotameters Total Flow	kg/hr	24.09	24.09	24.88	24.88	24.88	24.88	25.64	24.88
Purge % Flow	%	22.00	24.00	24.00	24.00	18.00		18.00	18.00
Purge Pressure	psig	14.50	14.50	14.50	14.00	14.50		14.30	14.50
Purge Flow	kg/hr	7.40	8.07	8.07	8.00	6.06		6.03	6.06
Sum Total Flow	kg/hr	98.67	99.91	101.26	100.14	99.13	93.08	99.89	99.68
Combustor Bottom Temp.	C	835.00	846.00	819.00	812.00	819.00	817.00	811.00	800.00
Combustor Middle Temp.	C	851.00	867.00	840.00	833.00	839.00	840.00	831.00	820.00
Combustor Top Temp.	C	852.00	868.00	841.00	834.00	840.00	842.00	832.00	821.00
Standpipe Top Temp.	C	820.00	827.00	822.00	814.00	820.00	820.00	813.00	804.00
Standpipe Middle Temp.	C	756.00	684.00	724.00	721.00	723.00	734.00	723.00	710.00
Standpipe Bottom Temp.	C	769.00	725.00	732.00	728.00	732.00	744.00	730.00	717.00
CO2 Concentration	%	15.70	15.00	14.10	13.10	13.20	14.60	13.80	13.10
O2 Concentration	%	3.50	3.90	5.40	5.00	4.40	4.40	4.90	5.50
CO Concentration	ppm	234.00	208.00	188.00	172.00	167.00	181.00	169.00	160.00
SO2 Concentration	ppm	1800.00		1800.00	2000.00	1900.00	2000.00	2100.00	2300.00
NOx Concentration	ppm	212.00	210.00	222.00	209.00	222.00	236.00	232.00	230.00

		2:00	3:00	4:00	5:00	6:00	7:00	8:00	9:00
Solids Feedrate	kg/hr		10.80	9.00	9.60	11.00	10.00	9.80	10.00
Secondary Cyclone Sample	kg		0.00	0.00	0.00	0.30	0.00	0.00	0.00
Secondary Cyclone Discharge	kg/hr					3.60			
J-Valve Sample	kg		0.00	0.00	0.00	0.90	0.00	0.00	0.00
Baghouse Discharge	kg/hr		2.70	1.80	2.40	2.70	3.00	2.80	3.30
Standpipe Discharge	kg/hr		0.10	0.10	0.10	0.50	0.30	0.80	0.90
Heat Exchanger Inlet Temp.	C		795.00	829.00	833.00	836.00	841.00	833.00	835.00
Heat Exchanger Outlet Temp.	C		774.00	782.00	792.00	790.00	770.00	771.00	785.00
Recycle Rate	kg/hr			1214.00	1595.00	1299.00	762.00	883.00	1298.00
Primary Air Differential	in. H20		30.00	30.00	30.00	30.00	30.00	30.00	31.00
Primary Air Pressure	psig		4.40	3.80	3.60	3.90	3.70	3.80	3.50
Primary Air Feedrat	kg/hr		61.71	60.73	60.40	60.90	60.57	60.73	61.23
Secondary Air Differential	in. H20		6.00	6.00	6.00	6.00	6.00	6.00	7.00
Secondary Air Pressure	psig		1.10	0.80	0.60	1.00	1.00	0.80	1.00
Secondary Air Feedrate	kg/hr		10.15	10.05	9.98	10.11	10.11	10.05	10.92
Total Flow Differential	in. H20		25.00	24.50	25.00	25.50	24.50	25.00	26.00
Total Flow Pressure	psig		15.50	15.50	16.00	16.00	16.00	16.00	17.00
Total Flow Feedrate	kg/hr		105.51	104.45	106.38	107.44	105.31	106.38	110.24
Rotameter 1 % Flow	%		30.00	25.00	25.00	25.00	25.00	25.00	25.00
Rotameter 1 Pressure	psig		4.00	2.25	3.50	3.50	3.50	2.10	1.50
Rotameter 1 Flow	kg/hr		4.53	3.59	3.72	3.72	3.72	3.58	3.51
Rotameter 2 % Flow	%		30.00	25.00	25.00	25.00	30.00	25.00	25.00
Rotameter 2 Pressure	psig		4.00	3.50	3.50	4.50	3.50	2.25	1.50
Rotameter 2 Flow	kg/hr		4.53	3.72	3.72	3.82	4.47	3.59	3.51
Rotameter 3 % Flow	%		20.00	20.00	22.00	22.50	25.00	20.00	20.00
Rotameter 3 Pressure	psig		4.25	3.00	3.00	3.50	3.50	3.50	2.90
Rotameter 3 Flow	kg/hr		3.04	2.94	3.23	3.35	3.72	2.98	2.93
Rotameter 4 % Flow	%		63.00	63.00	65.00	64.00	64.00	64.00	66.00
Rotameter 4 Pressure	psig		2.80	2.50	2.60	2.90	2.80	2.80	3.10
Rotameter 4 Flow	kg/hr		9.20	9.12	9.43	9.37	9.34	9.34	9.72
Rotameter 5 % Flow	%		20.00	20.00	20.00	25.00	21.00	22.00	25.00
Rotameter 5 Pressure	psig		1.50	2.00	1.75	1.50	2.00	1.80	2.10
Rotameter 5 Flow	kg/hr		2.81	2.85	2.83	3.51	2.99	3.12	3.58
Rotameter 6 % Flow	%		12.00	12.00	12.00	12.00	12.00	12.00	12.00
Rotameter 6 Pressure	psig		6.00	6.00	6.00	6.00	6.00	6.00	6.10
Rotameter 6 Flow	kg/hr		1.90	1.90	1.90	1.90	1.90	1.90	1.91
Sum Rotameters 1 to 6	kg/hr		26.00	24.12	24.84	25.68	26.15	24.51	25.15
Rotameters Total Flow Diff.	in. H20		8.00	7.50	8.00	8.00	8.50	8.00	8.00
Rotameters Total Flow Pressure	psig		15.00	15.00	15.00	15.00	15.50	15.50	16.50
Rotameters Total Flow	kg/hr		24.88	24.09	24.88	24.88	25.86	25.09	25.50
Purge % Flow	%		18.00	17.00	17.00	16.00	16.00	18.00	18.00
Purge Pressure	psig		14.40	14.60	14.60	14.70	14.60	14.50	14.60
Purge Flow	kg/hr		6.04	5.73	5.73	5.40	5.39	6.06	6.07
Sum Total Flow	kg/hr		102.78	100.60	101.00	101.29	101.93	101.92	103.72
Combustor Bottom Temp.	C		799.00	834.00	838.00	839.00	837.00	828.00	841.00
Combustor Middle Temp.	C		817.00	856.00	858.00	859.00	859.00	850.00	863.00
Combustor Top Temp.	C		818.00	858.00	860.00	861.00	861.00	848.00	866.00
Standpipe Top Temp.	C		795.00	831.00	833.00	836.00	841.00	834.00	841.00
Standpipe Middle Temp.	C		726.00	758.00	755.00	759.00	753.00	765.00	781.00
Standpipe Bottom Temp.	C		736.00	766.00	762.00	765.00	758.00	749.00	754.00
CO2 Concentration	%		14.30	13.80	14.00	14.00	14.00	14.50	15.70
O2 Concentration	%		5.00	4.80	4.40	4.40	4.30	4.40	4.30
CO Concentration	ppm		193.00	114.00	166.00	128.00	143.00	146.00	102.00
SO2 Concentration	ppm			1750.00	2000.00	1850.00	2000.00	1800.00	1400.00
NOx Concentration	ppm		226.00	254.00	251.00	252.00	246.00	241.00	254.00

		10:00	11:00	12:00	13:00	14:00
Solids Feedrate	kg/hr	12.20	10.80	10.20	11.40	11.20
Secondary Cyclone Sample	kg	0.35	0.00	0.00	0.00	0.30
Secondary Cyclone Discharge	kg/hr	4.20				3.60
J-Valve Sample	kg	0.70	0.00	0.00	0.00	0.70
Baghouse Discharge	kg/hr	3.20	3.40	3.00	3.70	3.20
Standpipe Discharge	kg/hr	0.50	0.20	0.50	0.20	1.10
Heat Exchanger Inlet Temp.	C	837.00	840.00	845.00	852.00	859.00
Heat Exchanger Outlet Temp.	C	788.00	789.00	790.00	797.00	808.00
Recycle Rate	kg/hr	1441.00	1404.00	1311.00	1051.00	815.00
Primary Air Differential	in. H20	31.00	31.00	30.00	33.00	32.00
Primary Air Pressure	psig	3.40	3.50	4.00	3.80	3.50
Primary Air Feedrate	kg/hr	61.07	61.23	61.06	63.70	62.21
Secondary Air Differential	in. H20	6.50	7.00	7.00	7.50	7.00
Secondary Air Pressure	psig	0.75	1.10	1.20	1.80	0.70
Secondary Air Feedrate	kg/hr	10.44	10.96	10.99	11.59	10.82
Total Flow Differential	in. H20	25.00	26.00	26.00	25.00	26.00
Total Flow Pressure	psig	17.00	17.50	16.00	17.00	17.00
Total Flow Feedrate	kg/hr	108.10	111.11	108.49	108.10	110.24
Rotameter 1 % Flow	%	25.00	26.00	25.00	25.00	25.00
Rotameter 1 Pressure	psig	2.00	2.00	2.50	2.10	2.50
Rotameter 1 Flow	kg/hr	3.56	3.71	3.62	3.58	3.62
Rotameter 2 % Flow	%	25.00	25.00	25.00	27.00	25.00
Rotameter 2 Pressure	psig	3.50	5.50	6.50	1.90	5.00
Rotameter 2 Flow	kg/hr	3.72	3.92	4.02	3.84	3.87
Rotameter 3 % Flow	%	22.00	27.00	25.00	23.00	20.00
Rotameter 3 Pressure	psig	2.00	3.00	2.20	2.50	6.50
Rotameter 3 Flow	kg/hr	3.14	3.96	3.59	3.33	3.21
Rotameter 4 % Flow	%	65.00	67.00	66.00	68.00	65.00
Rotameter 4 Pressure	psig	3.50	3.10	3.30	3.30	3.50
Rotameter 4 Flow	kg/hr	9.67	9.86	9.77	10.07	9.67
Rotameter 5 % Flow	%	20.00	21.00	22.00	24.00	25.00
Rotameter 5 Pressure	psig	1.80	2.50	1.80	2.00	2.30
Rotameter 5 Flow	kg/hr	2.83	3.04	3.12	3.42	3.60
Rotameter 6 % Flow	%	12.00	12.00	12.00	12.00	12.00
Rotameter 6 Pressure	psig	6.00	6.00	6.00	5.50	6.00
Rotameter 6 Flow	kg/hr	1.90	1.90	1.90	1.88	1.90
Sum Rotameters 1 to 6	kg/hr	24.84	26.40	26.01	26.11	25.88
Rotameters Total Flow Diff.	in. H20	8.00	8.50	8.00	8.00	8.00
Rotameters Total Flow Pressure	psig	16.50	17.00	15.50	16.50	16.50
Rotameters Total Flow	kg/hr	25.50	26.49	25.09	25.50	25.50
Purge % Flow	%	17.00	18.00	18.00	18.00	18.00
Purge Pressure	psig	14.60	14.80	14.80	14.60	14.40
Purge Flow	kg/hr	5.73	6.09	6.09	6.07	6.04
Sum Total Flow	kg/hr	102.74	104.77	103.23	106.85	104.58
Combustor Bottom Temp.	C	821.00	826.00	837.00	849.00	868.00
Combustor Middle Temp.	C	847.00	853.00	858.00	878.00	892.00
Combustor Top Temp.	C	847.00	855.00	864.00	881.00	890.00
Standpipe Top Temp.	C	838.00	841.00	845.00	861.00	857.00
Standpipe Middle Temp.	C	746.00	741.00	763.00	769.00	778.00
Standpipe Bottom Temp.	C	761.00	748.00	765.00	782.00	789.00
CO2 Concentration	%	15.10	15.60	15.60	15.60	16.20
O2 Concentration	%	4.70	4.30	3.50	3.60	3.00
CO Concentration	ppm	109.00	101.00	119.00	93.00	121.00
SO2 Concentration	ppm	1250.00	1300.00	1350.00	1650.00	2250.00
NOx Concentration	ppm	227.00	210.00	178.00	232.00	207.00

FILE: BATK2T1
TASK no: II TEST no:1

INPUT STREAMS

SPECIES		COAL	LIMESTONE
H2O	%	0.48	0.00
Ash	%	15.54	57.26
Total Carbon	%	73.16	11.50
Organic Carbon	%	73.16	0.00
Inorganic Carbon	%	0.00	11.50
Sulphur	%	6.34	0.00
Calcium	%	0.00	38.33
Hydrogen	%	4.68	0.00
Nitrogen	%	0.78	0.00
Oxygen	%	0.00	46.00
Total Flow	kg/h	8.24	2.70

FLUE GAS ANALYSIS

CO2	%	14.55
O2	%	4.39
CO	ppm	150.70
SO2	ppm	1805.56
NOx	ppm	227.55
N2	%	80.84
MOLE wt. of flue gas		30.58

CHEMICAL ANALYSIS AND FLOW RATES OF THE OUTPUT STREAMS

SPECIES		22:00 29	06:00 31	04:00 33	AVG.
				Cyclone	
Ash	%	78.50	79.00	80.40	79.30
Total Carbon	%	16.91	17.77	16.14	16.94
Organic Carbon	%			15.30	16.06
Inorganic Carbon	%			0.84	0.88
Total Sulphur	%	5.65	5.46	5.66	5.59
Sulphate Sulphur	%			5.25	5.25
Total Calcium	%	22.70	20.36	20.59	21.22
Total Flow	kg/h				0.06
				J-Valve	
Ash	%	99.65	99.70	99.65	99.67
Total Carbon	%	1.82	0.85	0.91	1.19
Organic Carbon	%			0.38	0.50
Inorganic Carbon	%			0.53	0.70
Total Sulphur	%	10.40	10.82	10.93	10.72
Sulphate Sulphur	%			10.98	10.98
Total Calcium	%	26.52	24.11	27.87	26.17
Total Flow	kg/h				0.15
				Baghouse	
Ash	%	77.65	78.70	80.85	79.07
Total Carbon	%	18.09	16.81	16.22	17.04
Organic Carbon	%			14.79	15.54
Inorganic Carbon	%			1.43	1.50
Total Sulphur	%	5.28	5.62	5.53	5.48
Sulphate Sulphur	%			5.15	5.15
Total Calcium	%	20.41	29.25	21.84	23.83
Total Flow	kg/h				2.71
				Standpipe	
Ash	%	91.90	96.90	97.95	95.58
Total Carbon	%	6.38	3.05	2.69	4.04
Organic Carbon	%			1.88	2.82
Inorganic Carbon	%			0.80	1.20
Total Sulphur	%	9.27	10.26	10.69	10.07
Sulphate Sulphur	%			11.07	11.07
Total Calcium	%	25.87	27.01	27.64	26.84
Total Flow	kg/h				0.54
				Weepage *	
Ash	%	99.65	99.70	99.65	99.67
Total Carbon	%	1.82	0.85	0.91	1.19
Organic Carbon	%			0.38	0.50

Inorganic Carbon	%			0.53	0.70
Total Sulphur	%	10.40	10.82	10.93	10.72
Sulphate Sulphur	%			10.98	10.98
Total Calcium	%	26.52	24.11	27.87	26.17
Total Flow	kg/h				0.50

* Assumed to be the same as the J-Valve

OVERALL MASS BALANCE

Air Feedrate	kg/h	101.36
Nitrogen Feed Rate	kg/h	78.05
Mass flowrate of dry flue gas	kg/h	103.98
Mass flowrate of wet flue gas	kg/h	111.40
Total Input Streams	kg/h	112.30
Total Output Streams	kg/h	115.36
Overall Mass Balance Error	%	-2.73
Theoretical Air Required	kg/h	84.99
Excess Air	%	19.26

CARBON BALANCE

Organic Carbon Input	kg/h	6.03
Inorganic Carbon Input	kg/h	0.31
Organic Carbon Out in Solids	kg/h	0.45
Inorganic Carbon Out in Solids	kg/h	0.05
Carbon Out in Flue Gas	kg/h	5.94
Carbon Balance Error	%	-1.65
Carbon Utilization (Solid Basis)	%	92.55
Carbon Utilization (Gas Basis)	%	94.52

SULPHUR BALANCE

Sulphur Input	kg/h	0.52
Sulphur Out in Solids	kg/h	0.28
Sulphur Out in Flue Gas	kg/h	0.20
Sulphur Balance Error	%	9.60
Sulphur Capture	%	62.39

CALCIUM BALANCE

Calcium Input	kg/h	1.03
Calcium Out in Solids	kg/h	0.97
Ca/S Ratio		1.58
Calcium Balance Error	%	5.92

ASH BALANCE

Ash Input	kg/h	2.83
Ash due to Sulphur fixation (Gas)	kg/h	0.81
Ash due to Sulphur fixation (Solids)	kg/h	0.69
Total Ash Input (Gas Basis)	kg/h	3.64
Total Ash Input (Solids Basis)	kg/h	3.52
Total Ash Out	kg/h	3.35
Ash Balance Error (Gas Basis)	%	7.89
Ash Balance Error (Solids Basis)	%	4.60

SOLIDS FEED RATE, BED TEMPERATURE AND GAS COMPOSITION FOR TEST # II-1

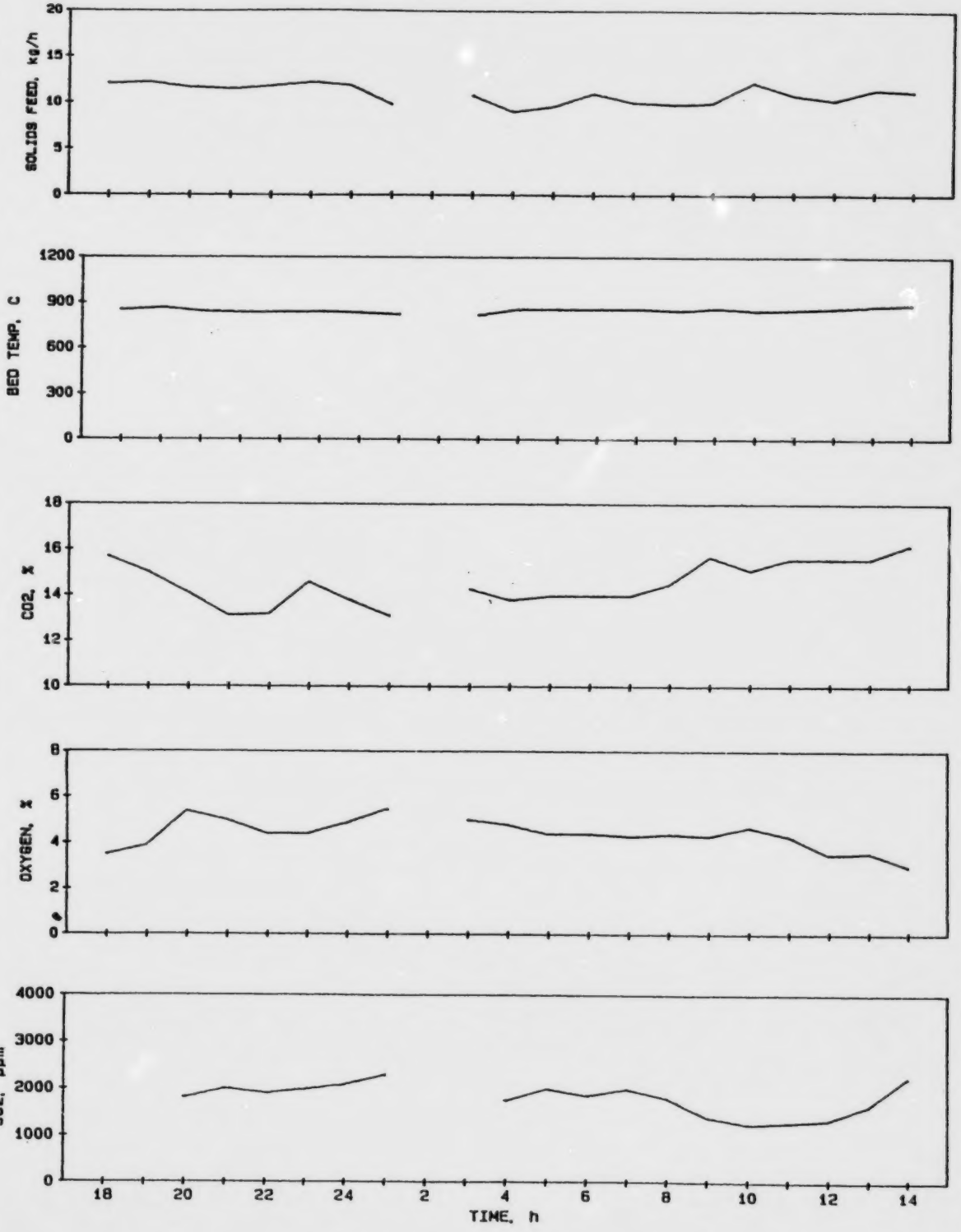


TABLE 1

Operating Conditions, Measures and Calculated Parameters

Task I

Parameter	Test 1	2	3
Operating Conditions:			
Feedrates, kg/h:			
Coal	8.48	8.29	8.32
Limestone	3.63	3.56	3.56
Air	103.28	102.14	103.21
Limestons Type and Size*	H-I	H-I	H-II
Ca/S Ratio	2:1	2:1	2:1
Air Distribution, %:			
Primary	56.4	57.9	57.2
Secondary	10.6	10.7	10.9
Others	33.0	31.3	31.9
Measured Parameters:			
Bed Temperature, °C	849	840	845
Discharge Rate, kg/h:			
Standpipe	0.15	0.34	0.52
Baghouse	3.82	4.05	3.27
Cyclone	0.10	----	----
Weepage	----	----	0.18
J-Valve	0.08	0.08	0.13
Cyclone Recycle,kg/h	4.15	----	----
Recirculation rate, kg/hr	1051	1197	1232
Gas Composition:			
CO_2%	14.0	14.2	15.1
O_2%	4.2	4.7	5.0
CO ppm	95	245	175
SO_2 ppm	1942	1964	1146
NO_x ppm	195	185	221
Calculated Parameters:			
Operating Velocity, m/s	7.1	6.9	7.0
Excess Air, %	20.4	21.8	22.6
Sulphur Capture, %	61.1	59.9	76.2
Carbon Utilization, %:			
Solid Basis	91.1	90.0	93.7
Gas Basis	89.7	93.7	102.0

* H = Havelock, Elm = Elmtree

TABLE 1 (cont'd)
Task II

Parameter	Test 1	2	3	4	6	7
Operating Conditions:						
Feedrates, kg/h:						
Coal	8.24	8.40	8.52	8.50	9.56	8.37
Limestone	2.70	2.76	5.49	3.65	4.10	3.59
Air	101.36	98.84	98.45	99.73	115.13	99.57
Air Distribution, %:						
Primary	59.1	61.2	61.3	60.6	60.9	60.8
Secondary	10.2	10.7	10.8	11.0	11.5	10.5
Others	30.7	28.1	28.0	28.4	27.7	28.7
Limestone Type and Size	H-II	H-II	H-II	H-III	H-II	Elm-I
Ca/S Ratio	1.5:1	1.5:1	2.5:1	2:1	2:1	2:1
Measured Parameters:						
Bed temperature, °C	852	857	851	851	856	851
Discharge Rate, kg/h:						
Standpipe	0.54	0.57	1.18	1.33	1.07	0.94
Baghouse	2.71	2.83	3.79	3.41	3.75	2.82
Cyclone	0.06	----	----	----	----	----
Weepage	----	0.47	0.56	0.57	0.20	0.54
J-Valve	0.15	0.16	0.17	0.18	0.15	0.18
Cyclone Recycle,kg/h	3.75	----	----	----	----	
Recirculation Rate, kg/hr	1023	896	838	950	1092	961
Gas Composition:						
CO_2%	14.6	14.8	14.5	14.9	13.9	14.4
O_2%	4.4	4.5	4.5	4.2	5.4	5.0
CO ppm	151	144	182	219	134	185
SO_2 ppm	1806	1987	1014	1314	1609	1840
NO_x ppm	228	241	250	240	247	228
Calculated Parameters:						
Operating Velocity, m/s	6.9	6.8	6.7	6.8	7.9	6.8
Excess Air, %	19.3	14.1	12.0	13.8	16.8	15.3
Sulphur Capture, %	62.4	60.3	80.2	73.9	67.1	62.7
Carbon Utilization, %:						
Solid Basis	92.6	92.2	92.2	91.7	92.6	91.7
Gas Basis	94.5	92.6	85.7	91.2	86.9	89.2

TABLE 1 (cont'd)

Task II

Parameter	Test 8	Test 9
Operating Conditions:		
Feedrates, kg/hr:		
Coal	8.32	8.41
Limestone	4.48	3.61
Air	99.47	98.93
Limestone Type and Size	ELM-II	ELM-II
Ca/S Ratio	2:1	2.5:1
Air Distribution, %:		
Primary	59.9	60.1
Secondary	10.4	10.5
Others	29.7	29.4
Measured parameters:		
Bed Temperature, °C	842	842
Discharge Rate, kg/hr:		
Standpipe	2.43	1.87
Baghouse	2.40	2.94
Cyclone	----	----
Weepage	0.48	0.47
J-Valve	0.27	0.21
Cyclone Recycle, kg/h	----	----
Recirculation Rate, kg/hr	958	1104
Gas Composition:		
CO_2%	14.3	14.0
O_2%	5.2	5.0
CO ppm	145	174
SO_2 ppm	832	1248
NO_x ppm	248	234
Calculated Parameters:		
Operating Velocity, m/s	6.8	6.7
Excess Air, %	15.9	14.0
Sulphur Capture, %	83.1	75.2
Carbon Utilization, %:		
Solid Basis	93.6	90.8
Gas Basis	88.0	85.3

TABLE 1 (cont'd)

Task III

Parameters	Test 1	2	3	5	6	7
Operating Conditions:						
Feedrates, kg/hr:						
Coal	9.29	8.35	8.28	8.89	7.97	8.29
Limestone	3.05	3.58	3.55	3.81	3.41	3.55
Air	100.93	100.72	94.61	102.49	92.96	99.12
Limestone Type and Size	H-II	H-II	H-II	H-II	H-II	H-II
Ca/S Ratio	2:1	2:1	2:1	2:1	2:1	2:1
Air Distribution, %:						
Primary	58.3	59.8	63.0	63.4	62.4	59.8
Secondary	----	----	----	----	----	----
Others	41.7	40.2	37.0	36.7	37.6	40.2
Measured Parameters:						
Bed Temperature, °C	847	817	846	805	907	848
Discharge Rate, Kg/hr:						
Standpipe	1.04	1.14	0.17	2.39	0.43	1.14
Baghouse	3.26	3.71	3.19	3.73	3.29	0.75
Cyclone	----	0.06	0.02	----	----	1.09
Weepage	0.50	0.68	0.56	0.43	0.42	0.50
J-Valve	0.13	0.15	0.32	0.27	0.13	0.74
Cyclone Recycle,kg/h	----	----	----	----	----	57.6
Recirculation Rate, kg/h	1233	2438	677	939	868	990
Gas Composition:						
CO_2%	15.1	12.9	14.6	13.7	14.3	14.9
O_2%	4.8	5.7	5.0	4.7	4.0	3.5
CO ppm	309	375	262	481	201	444
SO_2 ppm	1521	2169	1370	2544	1650	656
NO_x ppm	----	162	----	----	----	----
Calculated Parameters:						
Operating Velocity, m/s	6.9	6.7	6.5	6.7	6.7	6.8
Excess Air, %	16.8	29.7	22.8	23.9	25.4	28.5
Sulphur Capture, %	72.8	57.9	74.3	52.5	68.9	87.3
Carbon Utilization, %:						
Solid Basis	93.2	88.0	91.0	86.7	94.6	97.4
Gas Basis	99.3	91.3	98.6	94.8	96.9	104.4

TABLE 1 (cont'd)

Task III

Parameter	Test 4a	4b	4c
Operating Conditions:			
Feedrates, kg/hr:			
Coal	8.53	8.45	8.17
Limestone	3.93	0.10	3.53
Air	98.67	98.41	97.87
Limstone Type and Size	H-II	H-II	H-II
Ca/S Ratio	2.1:1	0:1	2.15:1
Air Distribution, %:			
Primary	60.5	61.0	61.0
Secondary	----	----	----
Others	39.5	39.0	39.0
Measured Parameters:			
Bed Temperature, °C	855	864	855
Discharge Rate, kg/hr:			
Standpipe	1.13	0.10	0.14
Baghouse	4.23	2.45	2.46
Cyclone	----	----	----
Weepage	0.55	0.75	0.29
J-Valve	1.38	0.48	0.23
Cyclone Recycle, kg/h	----	----	----
Recirculation Rate, kg/h	----	----	985
Gas Composition:			
CO_2%	15.6	13.9	15.3
O_2%	3.8	4.3	4.1
CO ppm	308	283	279
SO_2 ppm	1263	4025	1564
NO_x ppm	----	----	----
Calculated Parameters:			
Operating Velocity, m/s	6.8	6.8	6.7
Excess Air, %	24.3	25.2	28.8
Sulphur Capture, %	76.1	24.1	69.3
Carbon Utilization, %:			
Solid Basis	91.4	90.2	91.4
Gas Basis	106.9	101.6	108.1

TABLE 1 (cont'd)

Task IV

Parameter	Test 1	2	3	4	5
Operating Conditions:					
Feedrates, kg/hr:					
Coal	7.10	6.91	6.92	6.85	6.17
Limestone*	1.99	1.93	1.94	2.70	2.07
Shale	7.10	6.91	6.92	6.85	11.10
Air	97.08	98.15	97.77	99.41	99.87
Shale Size	I	I	II	II	II
Air Distribution, %:					
Primary	60.4	60.4	61.4	60.5	60.3
Secondary	9.8	10.1	10.2	9.7	9.7
Others	29.8	29.5	28.4	29.7	30.1
Measured Parameters:					
Bed Temperature, °C	838	848	850	847	851
Discharge Rate, kg/hr:					
Standpipe	4.91	1.78	2.76	2.58	3.81
Baghouse	2.56	5.21	4.85	5.51	7.10
Cyclone	1.42	0.32	----	----	----
Weepage	0.00	0.00	0.00	0.00	0.00
J-Valve	0.23	0.20	0.20	0.16	0.09
Cyclone Recycle,kg/h	31.93	----	----	----	----
Recirculation Rate,kg/h	----	958	935	839	1343
Gas Composition:					
CO_2%	13.4	13.6	12.7	13.7	14.6
O_2%	5.0	5.0	5.2	4.7	3.7
CO ppm	214	143	157	181	403
SO_2 ppm	1163	1602	1644	1348	1259
NO_x ppm	294	301	297	331	336
Calculated Parameters:					
Operating Velocity, m/s	6.6	6.7	6.7	6.8	6.8
Excess Air, %	18.9	23.5	22.8	26.2	24.8
Sulphur Capture, %	77.2	67.2	66.8	72.0	71.8
Carbon Utilization, %:					
Solid Basis	97.4	93.6	93.7	93.5	94.0
Gas Basis	88.0	93.1	85.5	93.4	100.3

* Havelock size II was used throughout Task IV.

TABLE 1 (cont'd)

Task V

Parameter	Test 1	2	3	4	5	6
Operating Conditions:						
Feedrates, kg/hr:						
Coal	8.52	8.47	8.38	8.20	8.40	8.57
Limestone	2.03	2.02	1.99	2.01	2.07	1.57
Air	94.20	97.66	95.53	95.83	96.07	95.70
Limestone Type+ and Size	H-II	Gln-18	Gln-35	Wnd-35	Wnd-18	Wnd-35
Ca/S Ratio	2:1	2:1	2:1	2:1	2:1	1.5:1
Air Distribution, %:						
Primary	60.8	59.5	59.8	59.8	59.5	59.0
Secondary	12.5	11.0	11.7	11.8	11.8	11.7
Others	26.8	29.6	28.6	28.4	28.7	29.3
Measured Parameters:						
Bed Temperature, °C	852	847	851	851	860	858
Discharge Rate, kg/hr:						
Standpipe	0.21	0.32	0.16	0.29	0.27	0.39
Baghouse (actual)*	1.58	3.24	1.40	1.69	1.35	6.61
Baghouse(corrected)*	2.61	2.60	2.38	2.37	2.5	1.98
Weepage	0.39	0.59	0.60	0.48	0.46	0.54
J-Valve	0.14	0.11	0.11	0.08	0.13	0.07
Cyclone Recycle,kg/h	----	----	----	----	----	----
Recirculation Rate, kg/h	863	895	1073	1162	1000	1087
Gas Composition:						
CO_2%	13.3	14.1	14.2	14.1	14.6	14.1
O_2%	5.3	4.8	4.8	4.5	4.4	4.9
CO ppm	209	219	209	190	197	175
SO_2 ppm	407	422	384	216	319	538
NO_x ppm	183	154	150	191	177	172
Calculated Parameters:						
Operating Velocity, m/s	6.5	6.7	6.5	6.6	6.6	6.6
Excess Air, %	12.7	17.7	16.4	19.3	16.8	14.0
Sulphur Capture, %	86.8	85.6	87.1	92.6	89.2	82.2
Carbon Utilization, %:						
Solid Basis(actual)*	95.7	87.3	96.3	94.9	95.9	81.1
Gas Basis(actual)*	80.4	89.7	89.2	90.2	91.8	91.0
Solid Basis(corrected)*	93.1	89.7	93.7	93.0	92.5	94.2
Gas Basis(corrected)*	80.6	89.6	89.4	90.3	91.8	88.8

\+ Gln = Glencoe, Wnd = Windsor

* The baghouse discharge was modified and the mass balances were re-calculated. More details are given in Section 4.3

4.1 Minto Coal and N.B. Limestones

Minto Coal and N.B. limestones were tested under a variety of operating conditions. Tasks I to III of this study, the effect of several operating parameters on sulphur capture have been identified. Those tested were as follows:

(a) Solids recirculation rate,
(b) Excess air,
(c) Secondary cyclone recycle,
(d) Ca/S ratio,
(e) Bed temperature,
(f) Operating velocity,
(g) Particle size,
(h) Limestone type.

The results for the first three Tasks are summarized in Table 2 (more details can be found in Table 1 and Appendix C). Each of the parameters mentioned above is discussed in detail below.

The particle size distribution of Havelock and Elmtree limestones are shown in Table 3 and Figures 5 to 8. Havelock I and Elmtree I were relatively fine with a mean particle size of 124 and 129 microns, respectively. The other limestones were much coarser with a mean particle size of no less than 305 μm. It is worthwhile noting that the 100 mesh fraction was removed from Havelock II and Elmtree I to produce Havelock III and Elmtree II, respectively. This fine fraction represents 7.3% in the Havelock limestone as compared to 40.6% in the Elmtree limestone.

TABLE 2

Summary of Test Conditions and Results
Minto Coal with Havelock & Elmtree Limestones

Test No.	Limestone Type	Limestone Size*	Ca:S Ratio	Operat. Vel. m/s	Bed. Temp. °C	Rec. Rate kg/h	Sec. Cycl.	Sul. Capt. %	Exc. Air %	Carbon Util. Gas %	Carbon Util. Solid %
Task I											
1	H	I	2:1	7.1	849	1051	yes	61.1	20.4	89.7	91.1
2	H	I	2:1	6.9	840	1197	no	59.9	21.8	93.7	90.0
3	H	II	2:1	7.0	845	1232	no	76.2	22.6	102.0	93.7
Task II											
1	H	II	1.5:1	6.9	852	1023	yes	62.4	19.3	94.5	92.6
2	H	II	1.5:1	6.8	857	896	no	60.3	14.1	92.6	92.2
3	H	II	2.5:1	6.7	851	838	no	80.2	12.0	85.7	92.2
4	H	III	2:1	6.8	851	950	no	73.9	13.8	91.2	91.7
6	H	II	2:1	7.9	856	1092	no	67.1	15.3	89.2	92.6
7	Elm	I	2:1	6.8	851	961	no	62.7	15.3	89.2	91.7
8	Elm	II	2.5:1	6.8	842	1104	no	83.1	15.9	88.0	93.6
9	Elm	II	2:1	6.7	842	958	no	75.2	14.0	85.3	90.8
Task III											
1	H	II	2:1	6.9	847	1233	no	72.8	16.8	99.3	93.2
2	H	II	2:1	6.7	817	2438	no	57.9	29.7	91.3	88.0
3	H	II	2:1	6.5	846	677	no	74.3	22.8	98.6	91.0
4a	H	II	2.1:1	6.8	855	—	no	76.1	24.3	106.9	91.4
4b	H	II	0:1	6.8	864	—	no	24.1	25.2	101.7	90.2
4c	H	II	2.15:1	6.7	855	985	no	69.3	28.8	108.1	91.4
5	H	II	2:1	6.7	805	939	no	52.5	23.9	94.8	86.7
6	H	II	2:1	6.7	907	868	no	68.9	25.4	96.9	94.6
7	H	II	2:1	6.8	848	990	yes	87.3	28.5	104.4	97.4

* See Table 3 for particle sizes.

TABLE 3

LIMESTONE PARTICLE SIZE DISTRIBUTION

MESH #	HAVELOCK			Elmtree	
	I	II	III	I	II
+14 (1.41 mm)	—	5.4	7.1	--	--
+16 (1.19 mm)	4.9	--	--	--	--
+18 (1.0 mm)	--	25.0	28.1	--	--
+20 (841 μm)	11.4	10.8	12.5	7.6	12.1
+25 (707 μm)	--	12.0	14.3	5.6	9.6
+30 (595 μm)	8.6	--	--	--	--
+45 (354 μm)	--	27.5	28.8	23.3	38.2
+50 (297 μm)	13.6	3.5	5.1	5.6	8.9
+60 (250 μm)	2.6	--	--	--	--
+100 (149 μm)	5.7	8.6	4.2	17.4	29.6
+120 (125 μm)	--	--	--	3.1	1.3
pan	53.2	7.3	0.2	37.5	0.4
Mean Particle Size, μm	124	400	660	129	366

FIGURE 5

LIMESTONE PARTICLE SIZE DISTRIBUTION

HAVELOCK I

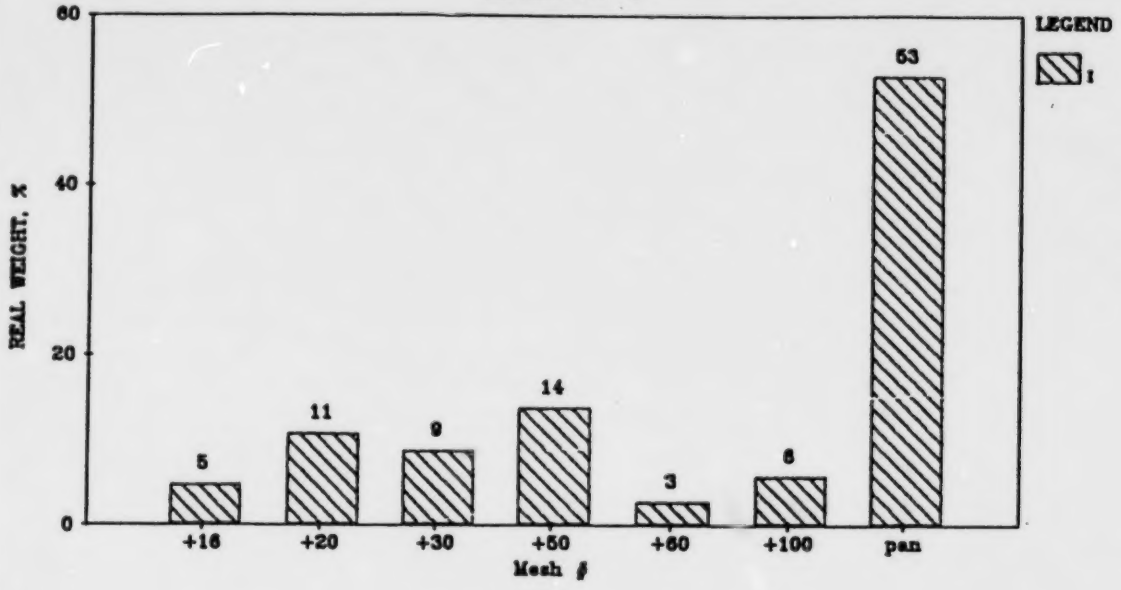


FIGURE 6

LIMESTONE PARTICLE SIZE DISTRIBUTION

HAVELOCK II & III

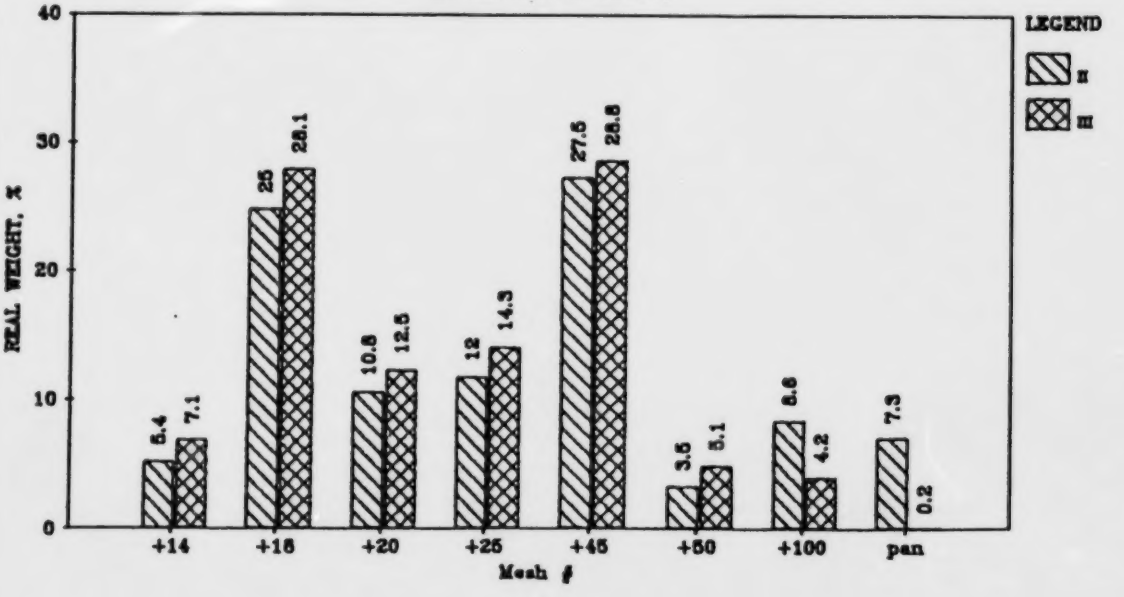


FIGURE 7

LIMESTONE PARTICLE SIZE DISTRIBUTION

ELMTREE I & II

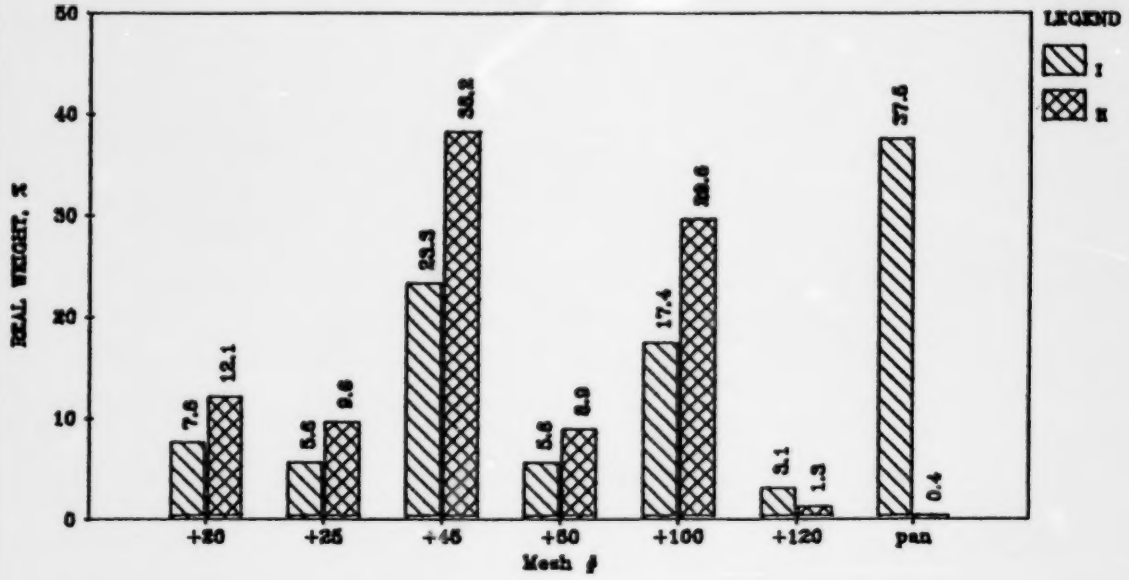
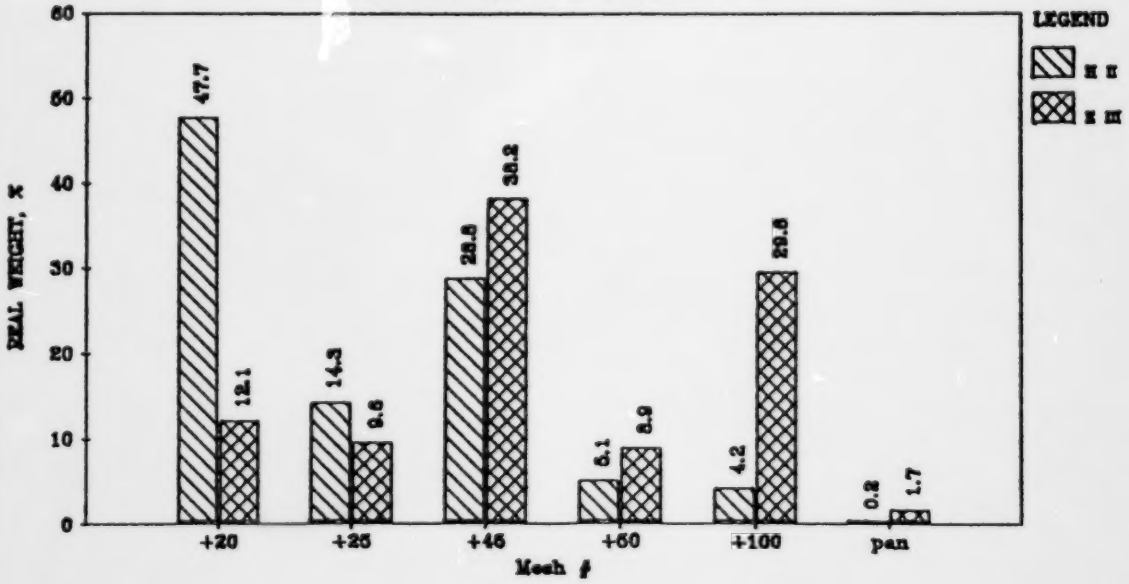


FIGURE 8

LIMESTONE PARTICLE SIZE DISTRIBUTION

HAVELOCK III & ELMTREE II



4.1.1 Effect of Recirculation Rate on Sulphur Capture

The rate of sulphur fixation in the combustor and primary cyclone does depend, among other parameters, on the population of calcium oxide particles which are normally available in appreciable quantities in the bed material. Increasing the rate of solid circulation increases the Ca/S ratio and, consequently, should be expected to lead to higher sulphur capture. For Minto coal, using 20% excess air and operating velocity of 7 m/s, recycling a 1000 kg/hr of bed material which is assumed to have only 10% of available calcium, gives a Ca/S ratio of approximately 144:1. This is much higher than the feed Ca/S ratios employed in this study which range between 1.5 to 2.5:1. The effect of Ca/S ratio on sulphur capture are discussed in Section 4.1.4. The results indicate that the feed Ca/S ratio, i.e., the fresh calcium oxide particles, are far more effective than the aged (less reactive) recycled particles. In order to assess the relative importance of the fresh and aged calcium particles an unsteady state test was carried out using a step change in the limestone feedrate. The test results and discussion are given in Section 4.5.

Figure 9 shows the effect of four recirculation rates on sulphur capture. For Tests III-3, I-3 and III-1 carried out at recirculation rates in the range of 677 to 1233 kg/hr, there is very little change in sulphur capture. While for test III-2 carried out a much higher recirculation rate of 2438 kg/hr, there is a significant drop in sulphur capture. Test III-2 results should be regarded with caution because the bed temperature was only 817°C. (Tests III-1, I-3 and III-3 were carried in the temperature range 845-847°C). The low bed temperature in Test III-2 was a direct result of recirculation of large quantities of relatively cold material into the combustor (heat is removed in the cooling section). Also, Test III-2 was the shortest recorded test lasting only for 8 hours because of difficulties in maintaining such high recirculation rate.

Further analysis of all experimental runs given in Table 2 are shown in Figure 10. All three straight lines are least squares fit for the following cases:

FIGURE 9

EFFECT OF RECIRCULATION RATE ON SULPHUR CAPTURE

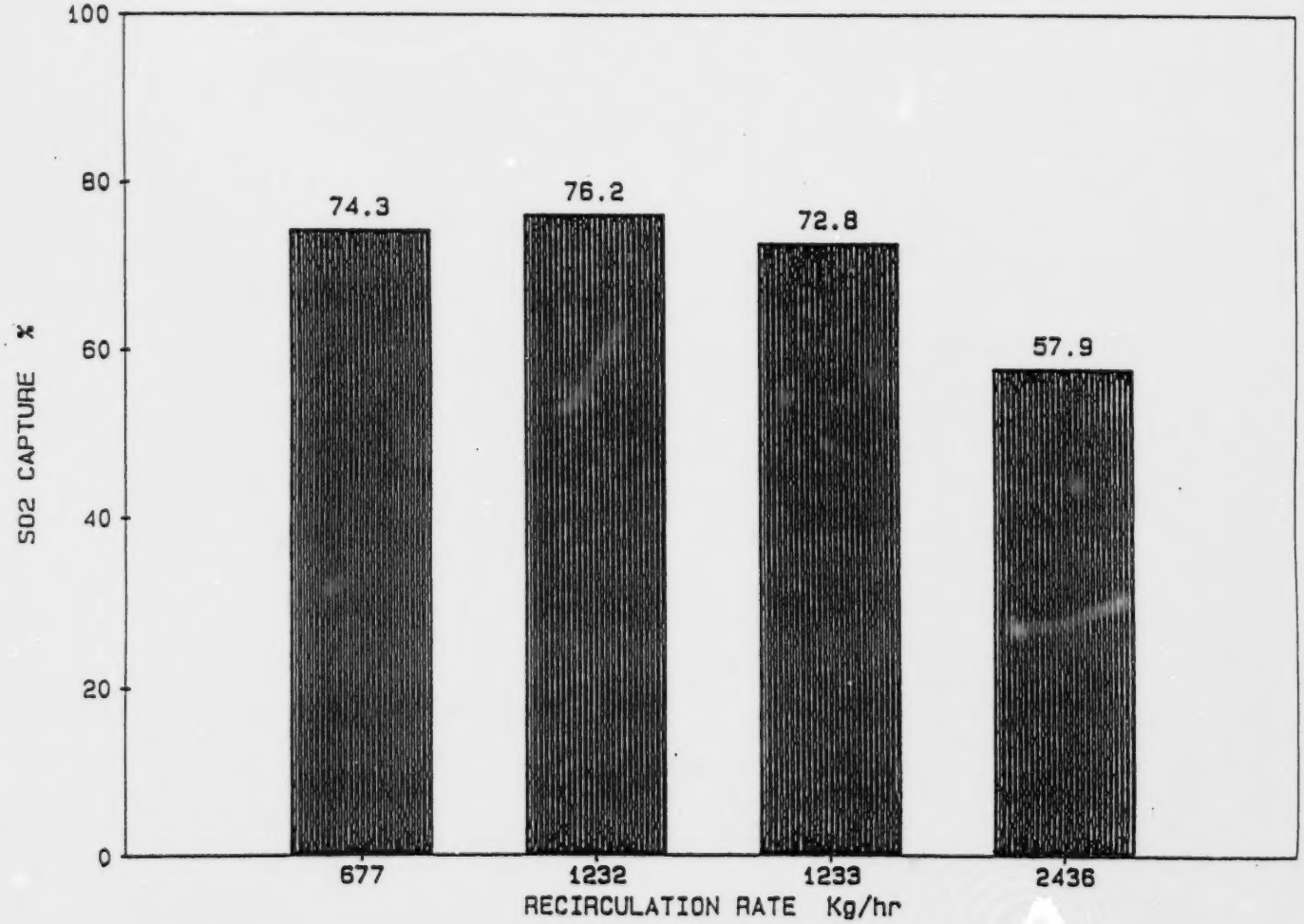
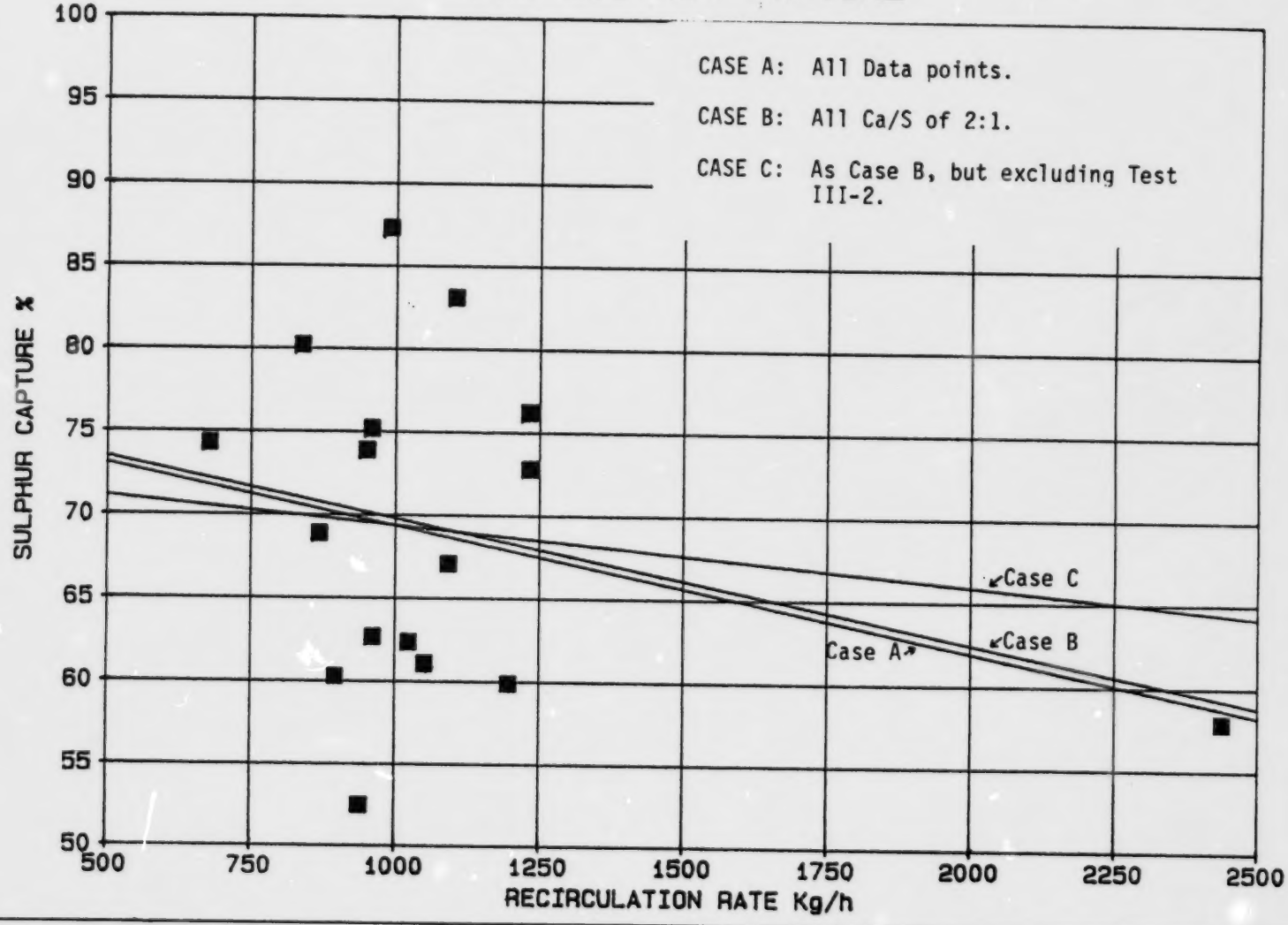


FIGURE 10

CORRELATION OF RECIRCULATION RATE AND SULPHUR CAPTURE



Case A: All the test results given in Table 2 except for Tests 4a, 4b and 4c.

Case B: All tests carried out at a Ca/S ratio of 2:1.

Case C: Same as Case B, but without Test III-2, which was carried out at a recirculation rate of 2438 kg/hr.

The correlation for all three cases are poor because there are wide differences it the operating condition, other than the recirculation rate. These parameters include limestone type, particle size, bed temperature, etc., all of which could have a significant impact on sulphur capture. Figure 10 does show, however, that increasing the recirculation rate leads to a slight decrease in sulphur capture (all three lines have a slightly negative slope). The impact, however, is very small particularly for Case C.

The main conclusion from the available results is that excessive recirculation rates do not improve sulphur capture. In fact it may result in loss of combustion efficiency, increase in bed pressure drop, i.e. high fan power requirement and potential erosion problems. On the other hand, loss of recirculation could lead to poor mixing, low heat transfer rates and consequently high bed temperature, all of which could lead to poor sulphur capture. On occasion, we have experienced loss of recirculation and it led to surges in bed temperature and a dramatic reduction in sulphur capture.

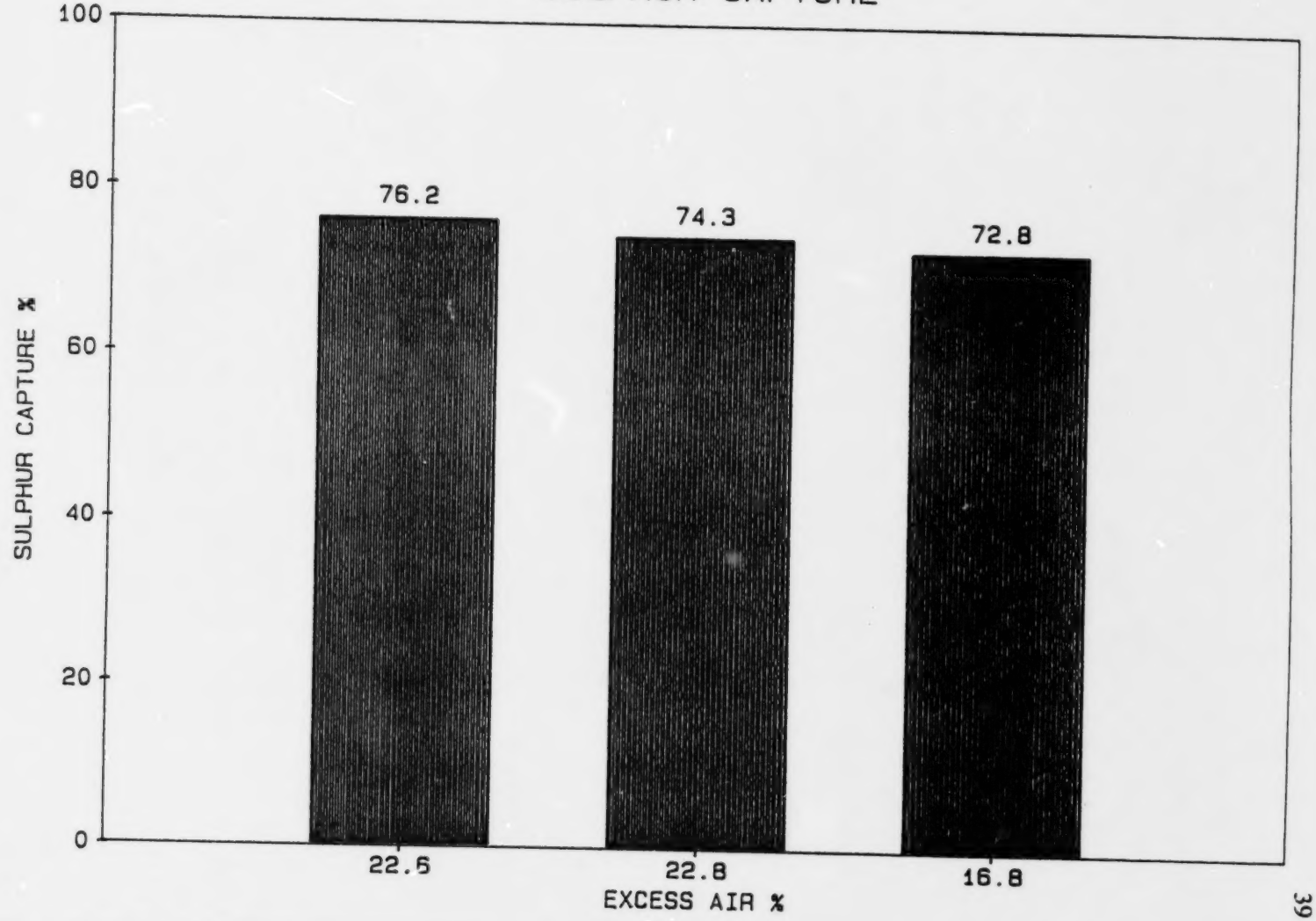
Based on these results, solid recirculation rate can be considered as a minor factor with regard to its effect on sulphur capture and will be regarded as such in subsequent data analysis.

4.1.2 Effect of Excess Air on Sulphur Capture

The effect of varying the excess air on sulphur capture can be illustrated by comparing the sulphur capture for Tests I-3, II-3, and III-1 carried out at an excess air levels of 22.6, 22.8 and 16.8%, respectively. The sulphur capture results are shown in Figure 11.

FIGURE 11

EFECT OF EXCESS AIR ON SULPHUR CAPTURE



The results indicate that within the range of the test conditions, the impact of varying excess air on sulphur capture are very small.

4.1.3 Effect of Secondary Cyclone Recycle on Sulphur Capture

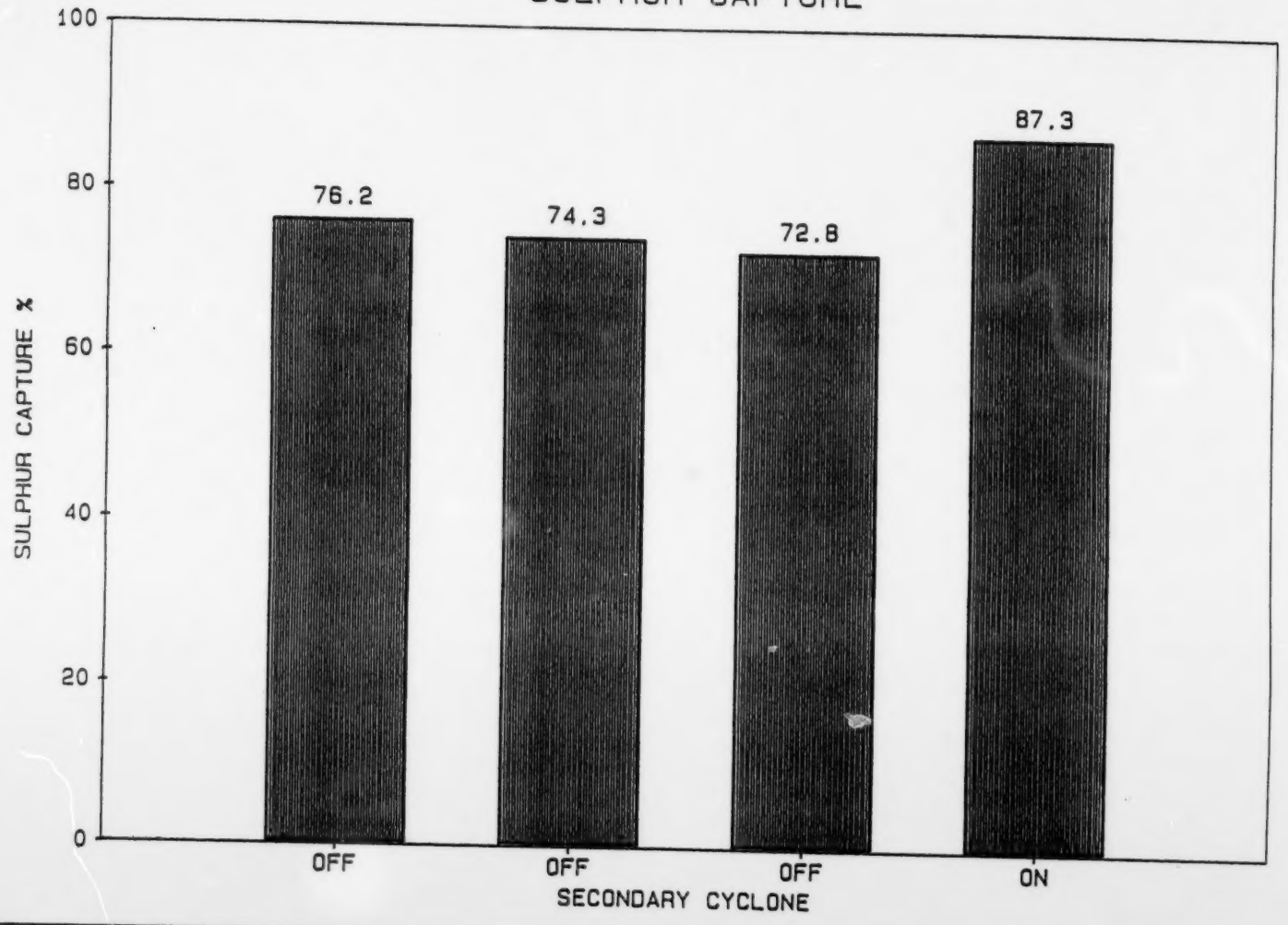
The effect of recycling the secondary cyclone catch back into the combustor has been evaluated in three different tests. The first two were carried out using a J-valve arrangement to provide the necessary seal to overcome the combustor backpressure as well as the pressure drop in the transportation line. This arrangement has been operated successfully under cold conditions. Tests 1 and 2 in the first two tasks were carried out under similar operating conditions with and without secondary cyclone recycle. As can be seen from Table 2, there is little difference in the sulphur capture. The reason for this negative result is simply that the dust was either recycled back at a very low rate or not at all. Evidence in support of this argument is given below.

In test III-7, the solid recycle system was modified to a lock-hopper arrangement with two 1.5" pneumatically operated valves acting as actuators. Also an observation window was included in the transport line. With this arrangement the dust recycled rate was measured using the procedure described in Section 3.0, at 57.6 kg/hr, see Table 1. This is compared with 4.15 and 3.75 kg/hr for tests I-1 and II-1, respectively. Note that for those two tests the measured rate of recycle is the same order of magnitude as the baghouse discharge. Therefore, it is likely that there was no recycle at all and the measured amount is simply the amount of dust which would have been normally collected in the baghouse had the stand-by cyclone not been in line. Based on this argument, Tests I-1, II-1 can be considered as no secondary cyclone recycle and will be treated as such in subsequent data analysis.

The sulphur capture achieved in test III-7 is 87.3%. This is much higher than that for tests I-3, III-1, III-3 as shown in Figure 12. The average sulphur capture for the latter three tests is 74.4%. Comparing this value with the 87.3% capture achieved in test III-7, corresponds to a 17% improvement in sulphur capture.

FIGURE 12

EFFECT OF SECONDARY CYCLONE ON SULPHUR CAPTURE



as a direct result of dust recycling. The significance of such improvement may be better illustrated by comparing SO_2 emission for the above mentioned tests. As can be seen from Table 1, the concentration of SO_2 in the flue gases for tests III-7, I-3, III-1 and II-3 are 656, 1146, 1521 and 1370 ppm. This means that approximately half of the SO_2 emission can be eliminated by using dust recycling.

For Test III-7 the rate of recycle was 57.6 kg/hr. This, when compared with a total feed rate of solids (coal and limestone) of 11.84 kg/hr, gives a recycle to feed ratio of 4.9:1. This is certainly high and, probably, similar results can be obtained at a much lower recycle rate. Previous work carried out at RPC in Phase I shows that slightly lower rates of sulphur capture can be achieved at a recycle ratio in the order of 1:1. The need for optimizing the rate of recycle is quite apparent, however, it was outside the scope of the present investigation.

4.1.4 Effect of Ca/S Ratio on Sulphur Capture

For Havelock limestone, the effect of varying Ca/S ratio is shown in Figure 13. The results shown are those for Test II-3 (2.5:1), Tests I-3, III-1 and III-3 (2:1) and Test II-1 and II-2 (1.5:1). As expected increasing Ca/S ratio leads to higher sulphur capture, however, as Ca/S ratio increases, calcium utilization decreases. For the three Ca/S ratios given above, the corresponding average calcium utilization figures are 32.1, 37.2* and 40.9%.

A similar trend was observed with Elmtree limestone. Tests II-8 and II-9 (see Table 2), were carried out at a Ca/S ratio of 2.5:1 and 2:1, respectively. The corresponding sulphur capture were 83.1 and 75.2%.

For both limestones increasing Ca/S ratios leads to a higher but diminishing increase in sulphur capture and a decrease in calcium utilization.

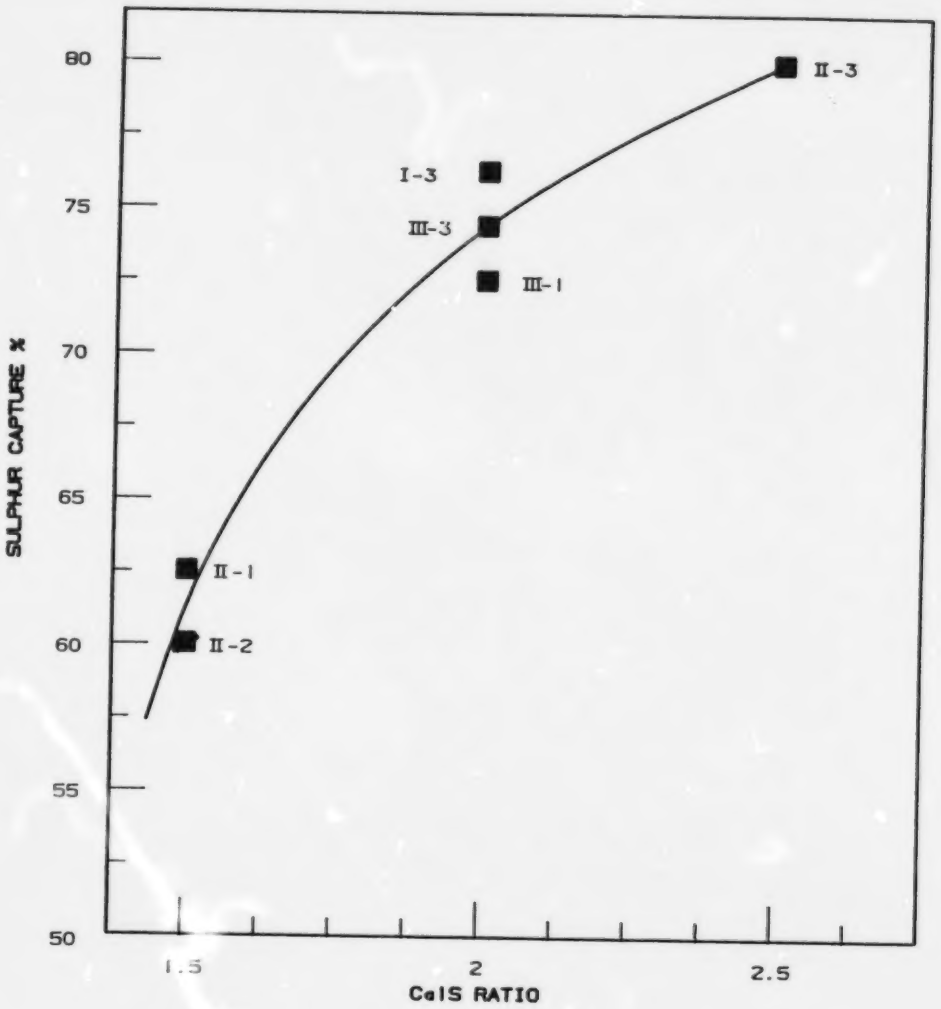


FIGURE 13
EFFECT OF CaIS RATIO ON SULPHURE CAPTURE

4.1.5 Effect of Bed Temperature Sulphur Capture

The results shown in Figure 14 are for Tests III-5, III-2, I-3, III-3, III-1 and III-6, carried out at bed temperatures of 805, 817, 845, 846, 847 and 907°C, respectively. As can be seen from Figure 14. Sulphur capture peaks at approximately 850°C. This is consistent with observations made by others, working on bubbling beds.

Temperatures lower than the optimum leads to a significant drop in sulphur capture. The results shown in Table 4 suggest that the low sulphur capture at 805°C is probably due to a reduction in the degree of limestone conversion into calcium oxide. Low temperature also results in a reduction in the rate of reaction between calcium oxide and SO_2. On the other hand, the reduction in sulphur capture at 907°C is probably due to limestone deactivation.

4.1.6 Effect of the Operating Velocity on Sulphur Capture

The operating velocity is considered as the superficial gas velocity based on total gas flowrate. The effect of varying the operating velocity on sulphur capture is illustrated in Figure 15. The results shown are for tests II-6, I-3, III-1, III-3, carried out at an operating velocity of 7.9, 7.0, 6.9 and 6.5 m/s, respectively. The results clearly show there is a decline in sulphur capture at the high operating velocity of 7.9 m/s. This can be attributed to a reduction in gas residence time in the combustor. However, the high gas velocity could also lead to high elutriation rate of limestone fines and consequently a decline in sulphur capture.

An attempt was made to conduct an additional test at an operating velocity of 5.75 m/s, however the test was abandoned due to severe slugging which led to frequent blockage of the feed line. The primary air represented 60% of the total air supply and the balance was distributed between the secondary and other air streams. This means the velocity at the bottom of the combustor was as low as 3.5 m/s. With a different air distribution and possibly finer feed (coal and limestone) it may be possible to operate at such low velocities.

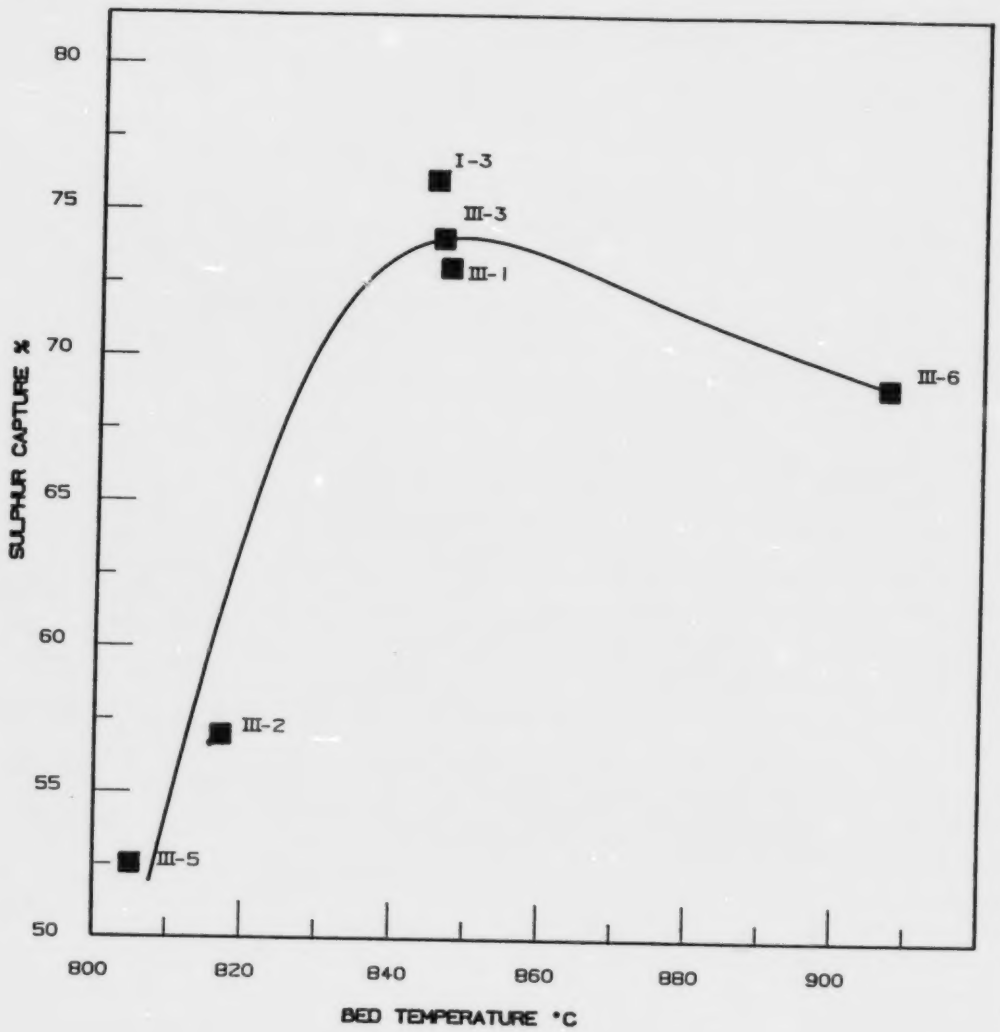


FIGURE 14
EFFECT OF BED TEMPERATURE ON SULPHURE CAPTURE

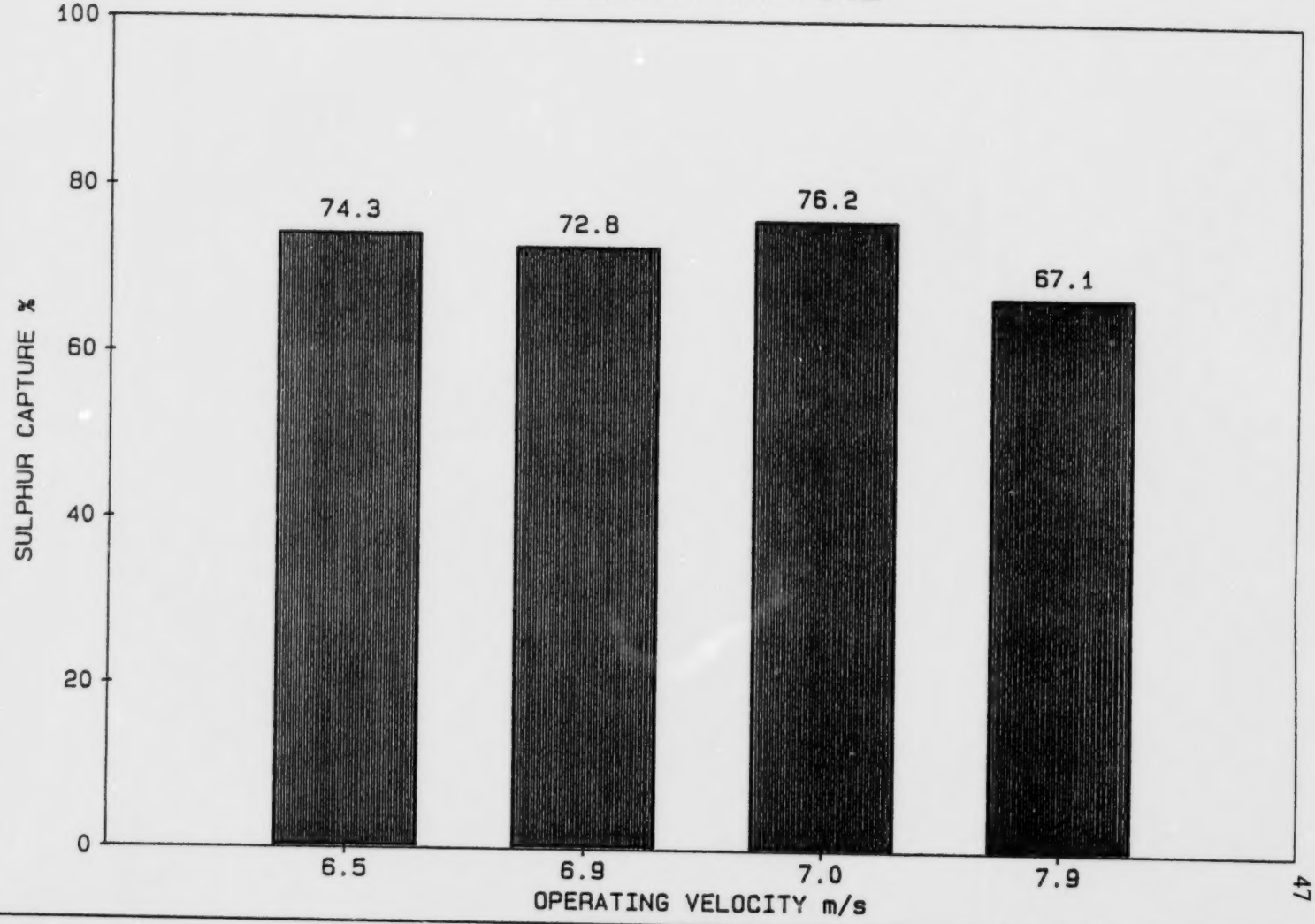
TABLE 4

CARBON CONTENT OF SELECTED BAGHOUSE PRODUCTS

Parameter	Test No. III-5	III-3	III-6
Combustor Temperature, °C	805	846	907
Carbon Content, wt%			
Total	23.08	15.55	8.98
Organic	19.99	14.63	8.15
Inorganic	3.09	0.91	0.83

FIGURE 15

EFFECT OF OPERATING VELOCITY ON SULPHUR CAPTURE



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4.1.7 Effect of Limestone Particle Size of Sulphur Capture

Tests I-2, I-3* and II-4 were carried out using three different sizes of Havelock limestone (see Table 3 and Figures 5 and 6). The degree of sulphur capture for these three tests are shown in Figure 16. The corresponding tests for Elmtree limestone were tests II-7 and II-9 and the results are shown in Figure 17.

For both limestones, Size I, gave low sulphur capture, 59.9% and 62.7% for Havelock and Elmtree limestones, respectively. The best sulphur capture results were achieved using Size II. These results suggest that very fine limestone particles are elutriated from the combustor before attaining high degree of sulphation. The vast majority of such particles would leave the system with the gas stream and would be collected as baghouse discharge. Very coarse limestone particles could also leave the system before reaching a high degree of sulphation. In this case, however, the material would report as bed discharge or as weepage through the distributor plate.

As mentioned earlier Havelock III and Elmtree II were prepared by screening out all -100 mesh particles from Havelock II and Elmtree I, respectively. In the case of Havelock limestone the impact of screening -100 mesh fraction on sulphur capture is very small. This could be due to the small percentage of -100 particles in Havelock II (see Table 3). Also, overscreening of Havelock II could be a factor. Table 3 shows that Havelock III, in comparison with Havelock II, has a large reduction not only in the -100 mesh size particles, but also in the 50-100 mesh size range, which is probably the most effective size fraction for sulphur capture. The loss of this size fraction may cause a reduction in sulphur capture despite the advantages gained by screening out the ultrafines of -100 mesh size.

* Test I-1 can be considered (assuming no secondary cyclone recycle) as a complement for Test I-2. Also, Tests III-1 and III-3 results can be used in conjunction with Test I-3.

FIGURE 16

EFFECT OF HAVELOCK PARTICLE SIZE ON SULPHUR CAPTURE

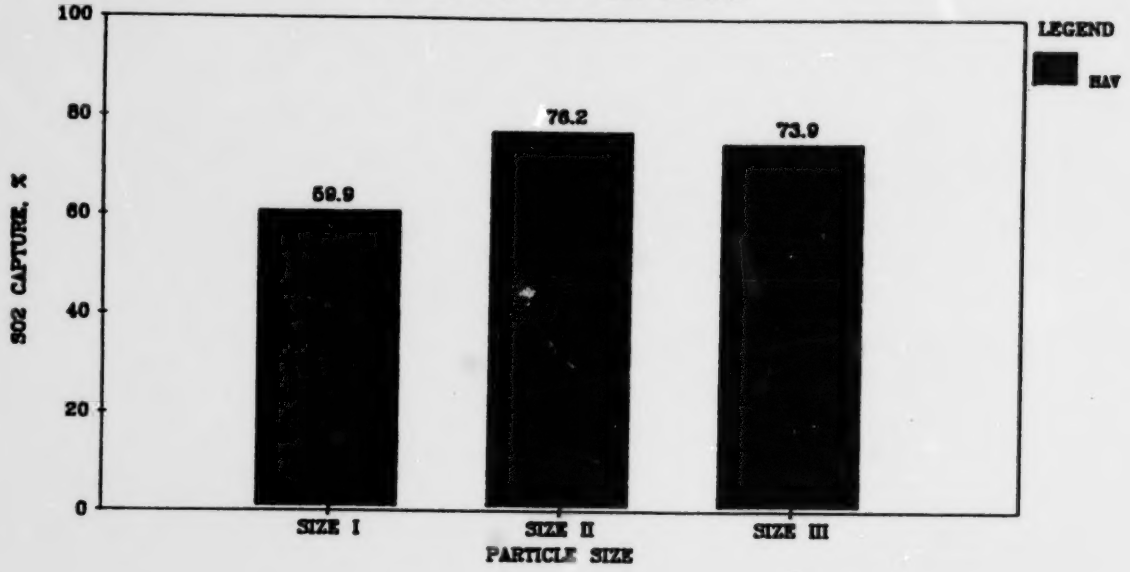
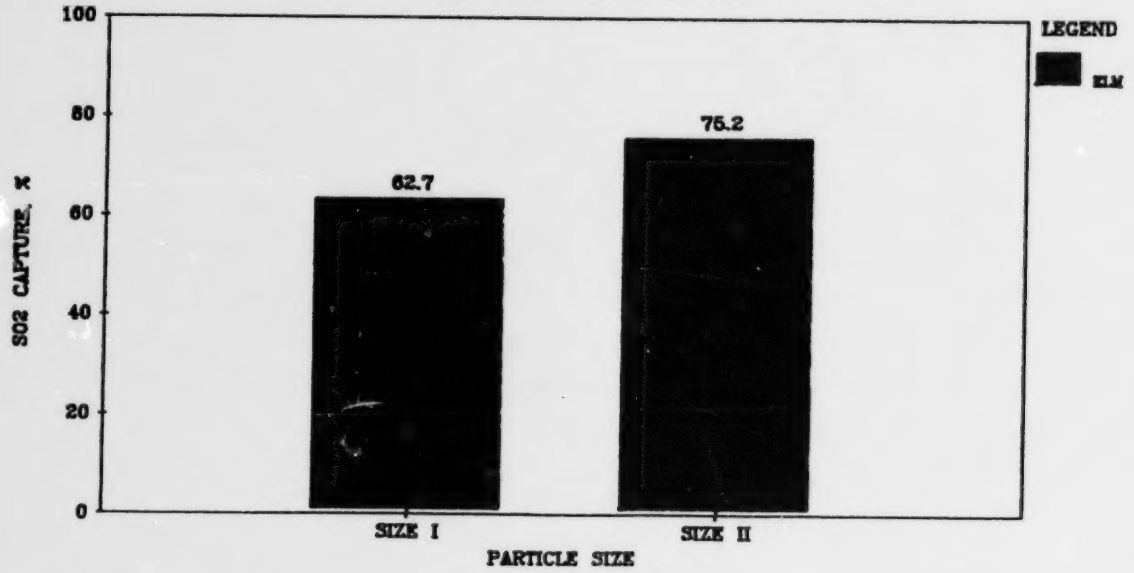


FIGURE 17

EFFECT OF ELMTREE PARTICLE SIZE ON SULPHUR CAPTURE



In the case of Elmtree, screening the ultrafine fraction led to a significant increase in sulphur capture. The change in the mean particle size was from 129 to 366 μm for Elmtree I and Elmtree II, respectively. The latter is probably closer to the optimum size which is a complex function of coal and limestone type and reactivity. Also, design parameters such as bed height, available residence time and cyclone efficiency can have a significant impact on the optimum size.

Microprobe and image analysis of some of the ash samples, see Appendix B, suggest that the most effective size range for the Havelock limestone is between 40 and 600 μm. This finding is in agreement with the results given above.

4.1.8 Effect of Limestone Type on Sulphur Capture

The Havelock and Elmtree limestones tested with Minto coal were of different particle size distribution as shown in Figures 5 to 8 and Table 3. However, there are sufficient similarities between the mean particle sizes of the two limestones to permit an evaluation of their performance.

The tests shown in Figures 16 and 17 were carried out at a Ca/S ration of 2:1. Comparing the results for size I for both limestones shows very little difference in their performance. The same conclusion can be reached by comparing size II results. Further evidence can be found in Table 2 by comparing the results for tests II-3 and II-8 carried at Ca/S ratio of 2.5:1. Sulphur capture for these tests were 80.2 and 83.1% for Havelock and Elmtree, respectively.

The above argument is based on the mean particle sizes, however, examining Table 3, shows that for both sizes, Elmtree limestone posses less ultrafines and no particles over 1 mm size. Judging from the strong particle size effect on sulphur capture, it may be inferred that Havelock is the better limestone. However, there are no concrete results to substantiate this conclusion. Therefore, further test work using closely matched particle size distribution would be necessary in order to rank the two limestones.

4.2 Minto Coal and N.B. Oil Shale and Limestones

The main objectives of this task were to assess the effect of shale particle size and coal/shale ratio on sulphur capture. The particle size distribution of the two shale sizes tested are given in Table 5 and Figure 18. The test results can be found in Table 1 and are summarized in Table 6.

4.2.1 Effect of Shale Particle Size on Sulphur Capture

The sulphur capture results for tests IV-2 and IV-3 are shown in Figure 19. The results indicate that changing, the mean particle size of shale from 301 to 461μm does not appear to have an impact on sulphur capture. This result should be regarded with caution because under the selected operating conditions*, 79% of the calcium requirement was supplied as limestone.

The calcium content of this particular batch of shale was relatively low at 2.86%. For a shale which contains 10% calcium, using similar set of operating conditions, the supplementary calcium requirement would represent only 26% of the total required.

4.2.2 Effect of Shale/Coal Ratio on Sulphur Capture

Test 4 and 5 were carried out at a Ca/S ratio of 2:1. This was intended in order to improve the low levels of sulphur capture of the previous tests. The shale/coal ratio varied from 1:1 for Test 4 to 1.8:1 for Test 5. The corresponding change in the fraction of the total calcium originated from the shales is 16% to 29% for Tests 4 and 5, respectively. The results shown in Figure 20 suggest that there is no significant difference between the reactivity (effectiveness) of oil shale or Havelock limestone.

*The feed consisted of equal amount of oil shale and coal. The Ca/S was kept low at 1.5:1 in order to minimize limestone addition.

TABLE 5

OIL SHALE PARTICLE SIZE DISTRIBUTION

Mesh #	Size I (fine fraction)	Size II (wide size)
+5 (4 mm)	--	5.1
+7 (2.83 mm)	--	16.1
+10 (2.0 mm)	6.8	20.0
+14 (1.41 mm)	23.8	15.2
+18 (1.0 mm)	15.5	9.6
+20 (841 μm)	6.1	3.7
+45 (354 μm)	22.8	14.7
+120 (125 μm)	12.4	8.5
pan	12.4	7.3
Mean particle size, μm	301	461

FIGURE 18

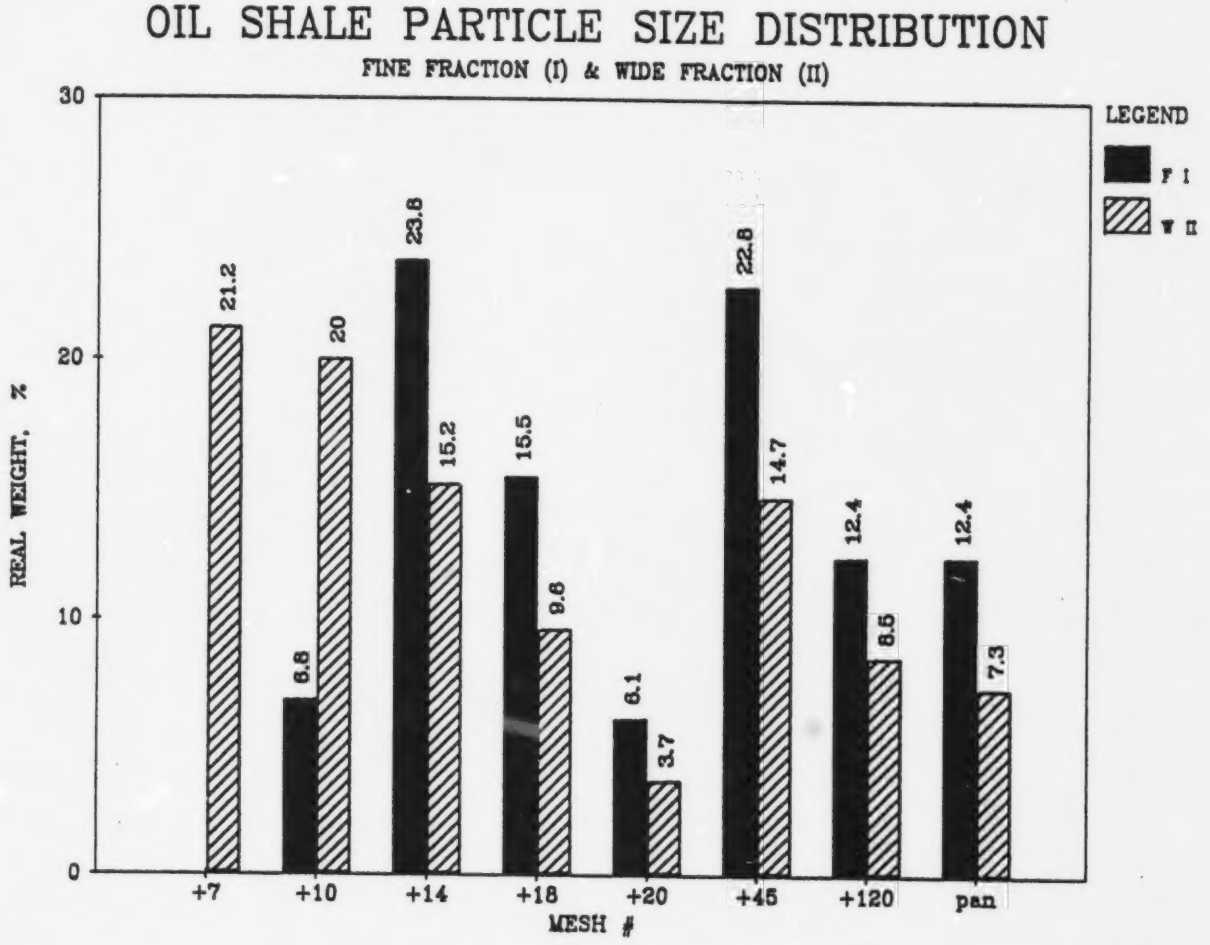

OIL SHALE PARTICLE SIZE DISTRIBUTION
FINE FRACTION (I) & WIDE FRACTION (II)
REAL WEIGHT, %
30
20
10
0
LEGEND
F I
W II
21.2
6.8
20
23.8
15.2
15.5
9.6
6.1
3.7
22.8
14.7
12.4
8.5
12.4
7.3
+7
+10
+14
+18
+20
+45
+120
pan
MESH #

TABLE 6

Summary of Test Conditions and Results
Minto Coal with Oil Shale and Havelock Limestone

Test No.	Shale/ Coal Ratio	Shale Coal	Ca:S* Ratio	Operat. Vel. m/s	Bed. Temp. °C	Rec. Rate kg/h	Sec. Cycl.	Sul. Capt. %	Exc. Air %	Carbon Util. Gas %	Carbon Util. Solid %
1	1:1	I	1.5:1	6.6	838	—	yes	77.2	18.9	88.0	97.4
2	1:1	I	1.5:1	6.7	848	958	no	67.2	23.5	93.1	93.6
3	1:1	II	1.5:1	6.7	850	935	no	66.8	22.8	85.5	93.7
4	1:1	II	2:1	6.8	847	839	no	72.0	26.2	93.9	93.5
5	1.8:1	Ii	2:1	6.8	851	1343	no	71.8	24.8	100.3	94.0

* Appropriate amounts of Havelock Size II were used in order to get the desired Ca/S ratio. Flowrates of coal, oil shale and limestone are given in Table 1.

FIGURE 19

EFFECT OF SHALE PARTICLE SIZE ON SULPHUR CAPTURE

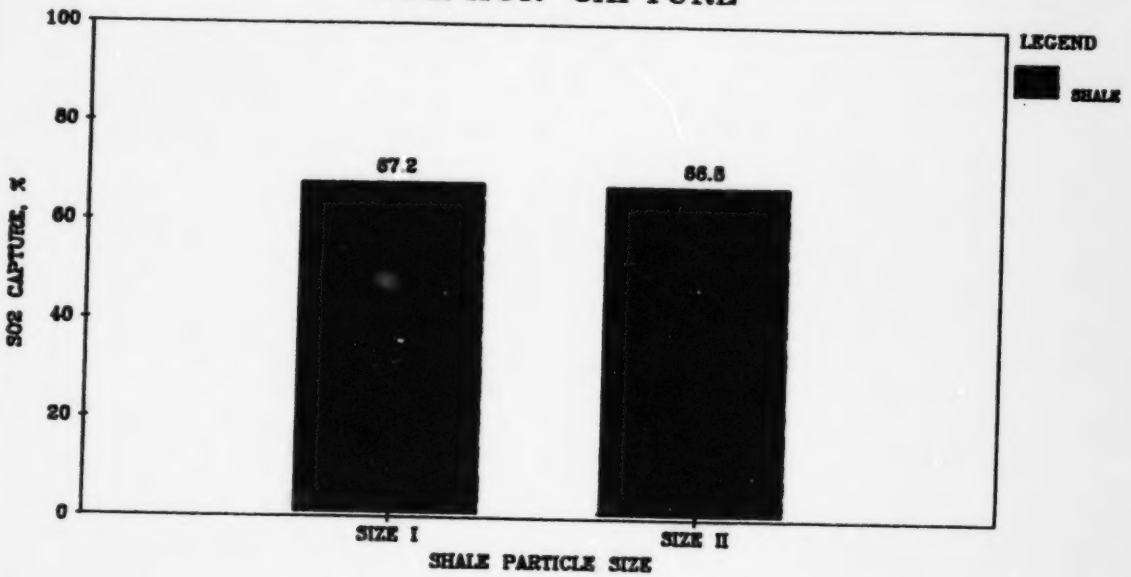
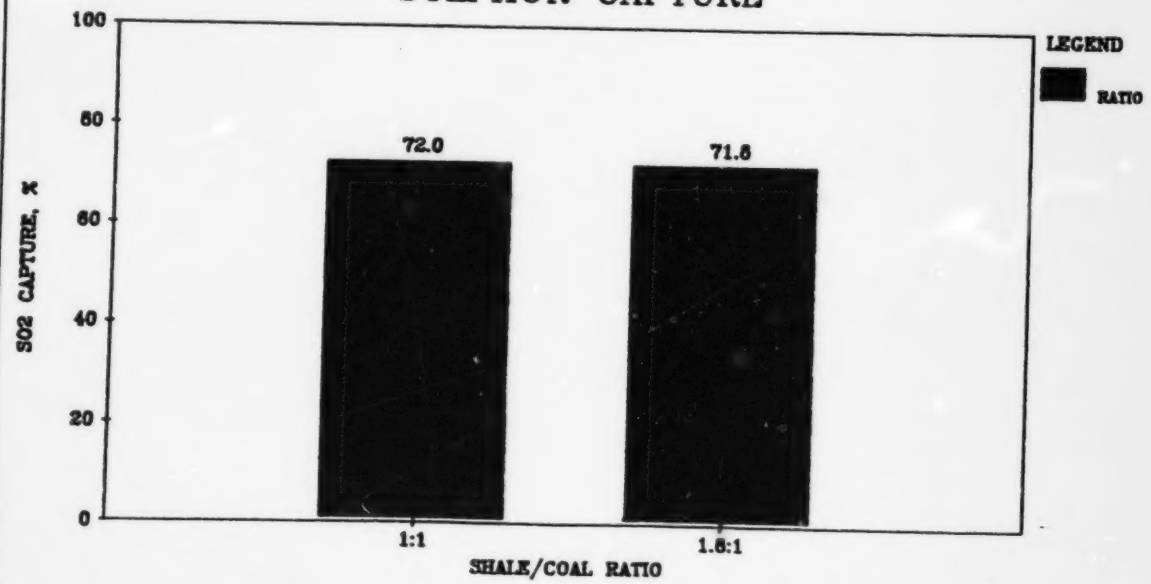


FIGURE 20

EFFECT OF SHALE/COAL RATIO ON SULPHUR CAPTURE



Two tests were attempted at an even higher oil shale to coal ratio of 2.2:1 and without any limestone addition. Such a high feedrate of shale led to difficulties with solid recycling from the standpipe to the combustor. This in turn caused wide fluctuations in bed temperature and poor sulphur capture. Poor recycle was probably caused by coarsening of the bed and reduction of the head in the standpipe (shale has a low bulk density). Also, the flaky nature of the shale may not be conducive to solids flow. It is important to note that feed of such high shale to coal ratio could be burned successfully by making some changes such as, smaller shale top size, larger bed inventory (provide more time for particle attrition) and a standpipe of a greater height.

4.2.3 Effect of Secondary Cyclone Recycle on Sulphur Capture

Earlier test results given in Section 4.1.3 have shown that dust recycle can lead to dramatic improvement in sulphur capture. The same conclusion can be reached by comparing the sulphur capture results for Tests 1 and 2 given in Table 6.

4.3 Prince Coal and Nova Scotia Limestones

In this task the Nova Scotia feedstocks were evaluated. Prince coal was burnt with Glencoe and Windsor limestones. Also, a base test was carried out using Havelock limestone. The two primary objectives of this task were to evaluate the effect of limestone type and particle size on sulphur capture.

4.3.1 Effect Limestone Particle Size on Sulphur Capture

The particle size distribution of the Nova Scotia limestones are shown in Table 7. Size II limestones were taken from the same batch as size I by removing particles coarser than 35 mesh. The particle size distribution of prince coal and Size I limestones are shown in Figures 21 and 22, respectively.

The test results for Nova Scotia feedstocks are summarized in Table 8. The results show that for both limestones removing the coarse fraction which is approximately 25% by weight led to small improvement in sulphur capture. The changes in sulphur capture were 85.6 to 87% and 89.2 to 92.6% for Glencoe and Windsor limestones, respectively. This improvement is probably not sufficient to justify using the finer limestone on a commercial scale.

The mean particle sizes of all the Nova Scotia limestones are fairly small and are comparable in value to Size I of Elmtree and Havelock limestones both of which gave poor sulphur capture. One of the probable causes for such difference in performance is the fact that the two N.B. limestones in question did have a wide particle size distribution and have a large fraction of very fine particle of less than 125 μm.

The results discussed above confirm the earlier conclusion given in Section 4.1.7, namely that the two extremes in the particle size spectrum are less effective than the middle portion. The mineralogical study, given in Appendix C, suggest the boundaries to be 600 to 40 μm.

TABLE 7
PARTICLE SIZE DISTRIBUTION OF NOVA SCOTIA LIMESTONES

	Glencoe		Windsor	
	I	II	I	II
+18 (1.0 mm)	0.1	----	0.1	----
+25 (707 μm)	9.4	----	10.6	----
+35 (500 μm)	14.0	0.1	15.3	0.1
+50 (297 μm)	19.1	25.0	25.9	35.0
+60 (250 μm)	8.7	11.4	8.8	11.9
+80 (177 μm)	10.9	14.2	9.4	12.7
+120 (125 μm)	11.0	14.4	11.4	15.4
+200 (74 μm)	13.1	17.1	9.6	13.0
pan	13.6	17.8	8.9	12.0
Particle Size, μm	136	110	170	135

FIGURE 21

PRINCE COAL PARTICLE SIZE DISTRIBUTION

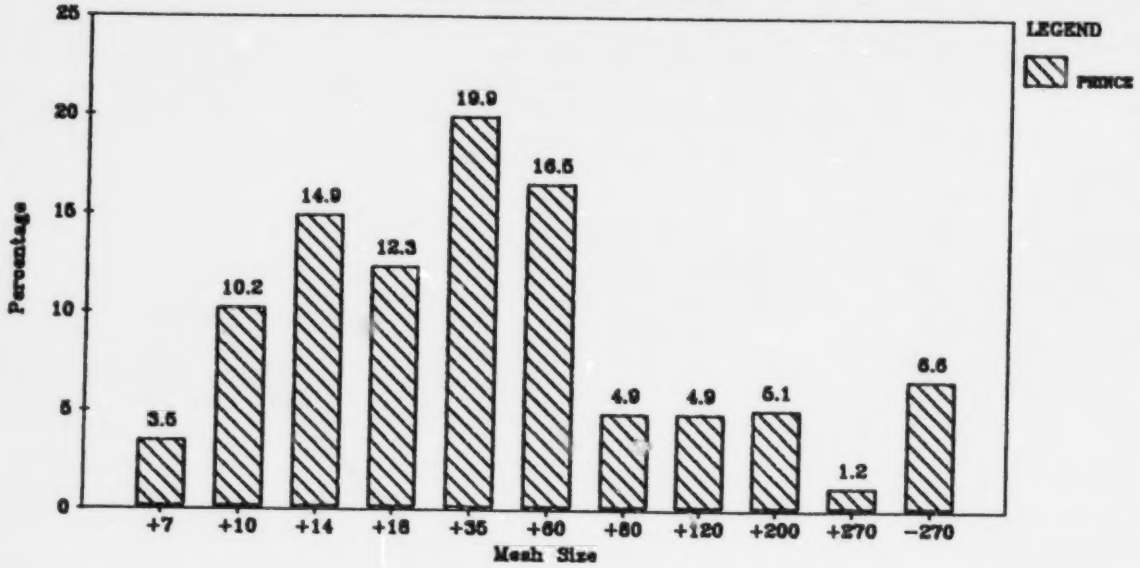


FIGURE 22

LIMESTONES PARTICLE SIZE DISTRIBUTION

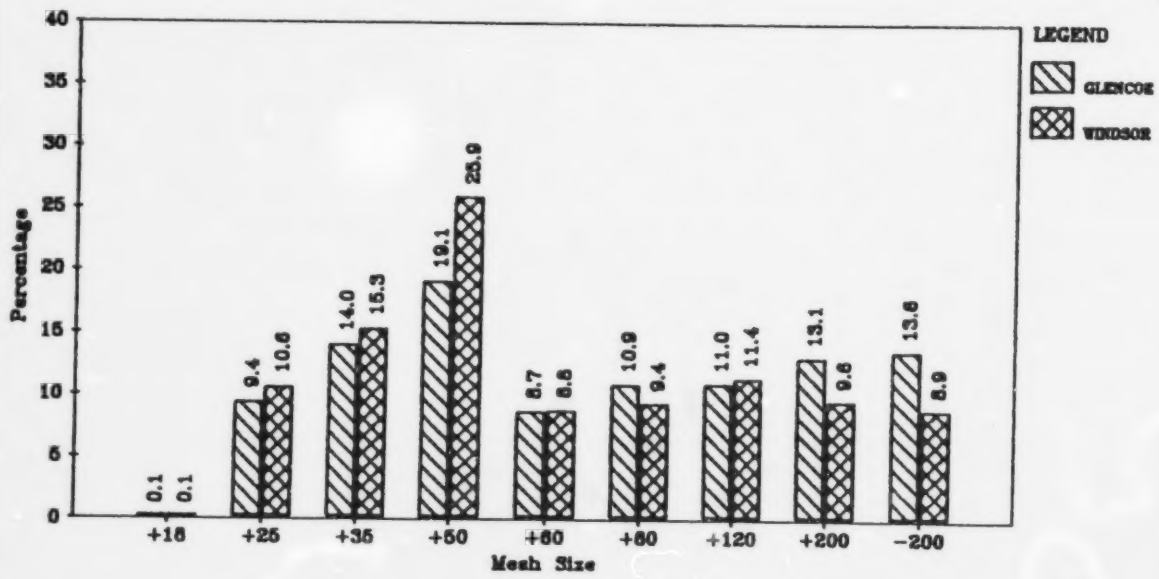


TABLE 8

Summary of Test Conditions and Results
Prince Coal with Havelock, Glencoe and Windsor Limestones

Test No.	Limestone Type	Size*	Ca:S Ratio	Operat. Vel. m/s	Bed Temp. °C	Rec. Rate kg/h	Sul. Capt. %	Exc. Air %	Carbon Util. Gas %	Carbon Util. Solids %
1	H	II	2:1	6.5	852	863	86.8	12.9	80.4(80.6)	95.7(93.1)
2	Gln	I	2:1	6.7	847	895	85.6	17.7	89.7(89.6)	87.6(89.7)
3	Gln	II	2:1	6.5	851	1073	87.0	16.4	89.2(89.4)	96.3(93.7)
4	Wnd	II	2:1	6.6	851	1162	92.6	19.3	90.2(90.3)	94.9(93.0)
5	Wnd	I	2:1	6.6	860	1000	89.2	16.8	91.8(91.9)	95.9(92.5)
6	Wnd	II	1.5:1	6.6	858	1087	82.2	14.0	91.0(88.8)	81.1(94.2)

* Values in parenthesis are for corrected baghouse discharge. See Section 4.4.

4.3.2 Effect of Limestone Type on Sulphur Capture

The results given in Table 8 shows that Windsor is more reactive than Glencoe. Also, the latter is hard crystallized limestone and it proved difficult to crush using our lab scale crushing facility.

In the base line test carried out using Havelock limestone Size II, a sulphur capture of 86.8%, has been achieved which is comparable to the Glencoe results. Using the same limestone with Minto coal, the results given in Table 2 show that the attainable sulphur capture is approximately 10% lower than this result. This is an interesting finding and it shows the impact of coal characteristics on sulphur capture. Minto coal has almost twice the sulphur content of Prince coal. Also, the coal ash itself can retain sulphur as well as enhance the reactivity of a particular limestone. Certain additives have been shown to enhance the capability of particular limestone to capture sulphur by up to 3 fold (2).

4.3.3 Effect of Ca/S Ratio on Sulphur Capture

The effect of Ca/S ratio on sulphur can be illustrated by comparing the results for Tests V-6 and V-4, carried out at a Ca/S ratio of 1.5:1 and 2:1, respectively. As can be seen from Table 8, increasing Ca/S ratio from 1.5:1 to 2:1, led to a significant increase in sulphur capture (82.2 to 92.6%, respectively). The concentration of SO_2 in the flue gases for Tests V-6 and V-4, as shown in Table 1, is 538 and 216 ppm, respectively. Therefore, SO_2 emission was reduced by 60% as a result of increasing Ca/S from 1.5:1 to 2:1.

These results are consistent with those for Minto coal given in Section 4.1.4, and as expected it clearly shows taht Ca/S ratio is among the significant parameters effecting sulphur capture.

4.4 Mass Balances

For each experimental run the overall mass and component balances were carried out and the results are given in Appendix C. The mass balance errors* were tabulated and are shown in Table 9 and 11.

A review of the results for Tasks I to IV, given in Table 9 and Figures 23-26, shows that the errors for the overall and carbon balances are extremely small and have a mean value of -1.2 and 0.1%, respectively. The standard deviation are also low indicating accurate mass balances.

The sulphur, calcium and ash mass balance errors shown in Table 9, although higher than those given above, are still low. It is important to note that the errors for the three components are similar in magnitude. Knowing that all of the calcium and 70 to 80% of the sulphur report in the solid (ash) streams, any improvement in the ash balance would give to a similar improvement in the other two components. The impact on the carbon balance should be negligible (most of the carbon burns away). However, the overall mass balance would be affected.

The ash losses can be contributed to several factors including experimental errors in measuring the solid input and output streams, the errors may indicate ash accumulation in the system, particularly in the baghouse very fine ash particles escaping the baghouse and analytical errors which can be caused by hydration of the solid samples leading to under estimation of the ash.

In Task V, the baghouse discharge was very erratic. In the previous 4 tasks using N.B. feedstocks, the baghouse performance was consistent with the hourly discharge within 10-20% of the mean value. The altered performance with Prince Coal could perhaps be due to the nature of the ash produced from the combustion of the N.S. feedstock under

* $$\% \text{ Error component x} = \left(1 - \frac{\text{component x out of the system}}{\text{component x into the system}}\right) \times 100$$

TABLE 9
MASS BALANCE ERRORS
FOR TASKS I TO IV

Test No.	Overall	Component Error, wt% Carbon	Sulphur	Calcium	Ash
Task I					
1	1.6	1.9	21.8	20.0	13.4
2	0.1	-3.1	16.4	8.5	7.4
3	-0.1	-7.9	29.4	12.1	10.5
Task II					
1	-2.7	-1.7	9.6	5.9	4.6
2	-3.3	-0.03	11.7	18.2	4.9
3	-2.2	7.0	-3.7	13.4	1.1
4	-3.4	0.9	-7.2	-6.7	-6.2
6	-2.5	6.1	3.9	15.9	6.4
7	-2.9	3.0	-3.6	26.3	10.0
8	-2.8	6.0	-11.9	15.5	-4.9
9	-3.0	6.0	-12.9	6.9	-4.4
Task III					
1	-4.4	-5.7	10.4	-20.5	5.6
2	-1.8	-2.5	17.3	-4.1	1.9
3	-2.9	-6.9	18.1	12.9	19.5
5	-3.3	-7.4	-13.7	-12.4	-5.2
6	-1.4	-2.0	6.2	0.1	11.2
7	-1.6	-6.7	10.7	6.4	14.7
Task IV					
1	1.2	9.4	-10.5	5.1	2.6
2	2.0	0.8	24.2	18.2	19.1
3	2.7	8.5	14.1	14.3	12.7
4	2.3	1.8	14.1	8.9	11.9
5	1.7	-6.0	3.7	-2.1	10.9
Mean	-1.2	0.1	6.7	7.4	6.7
Standard Deviation	2.2	5.5	12.8	11.3	7.6

FIGURE 23

MASS BALANCE ERRORS

TASK I

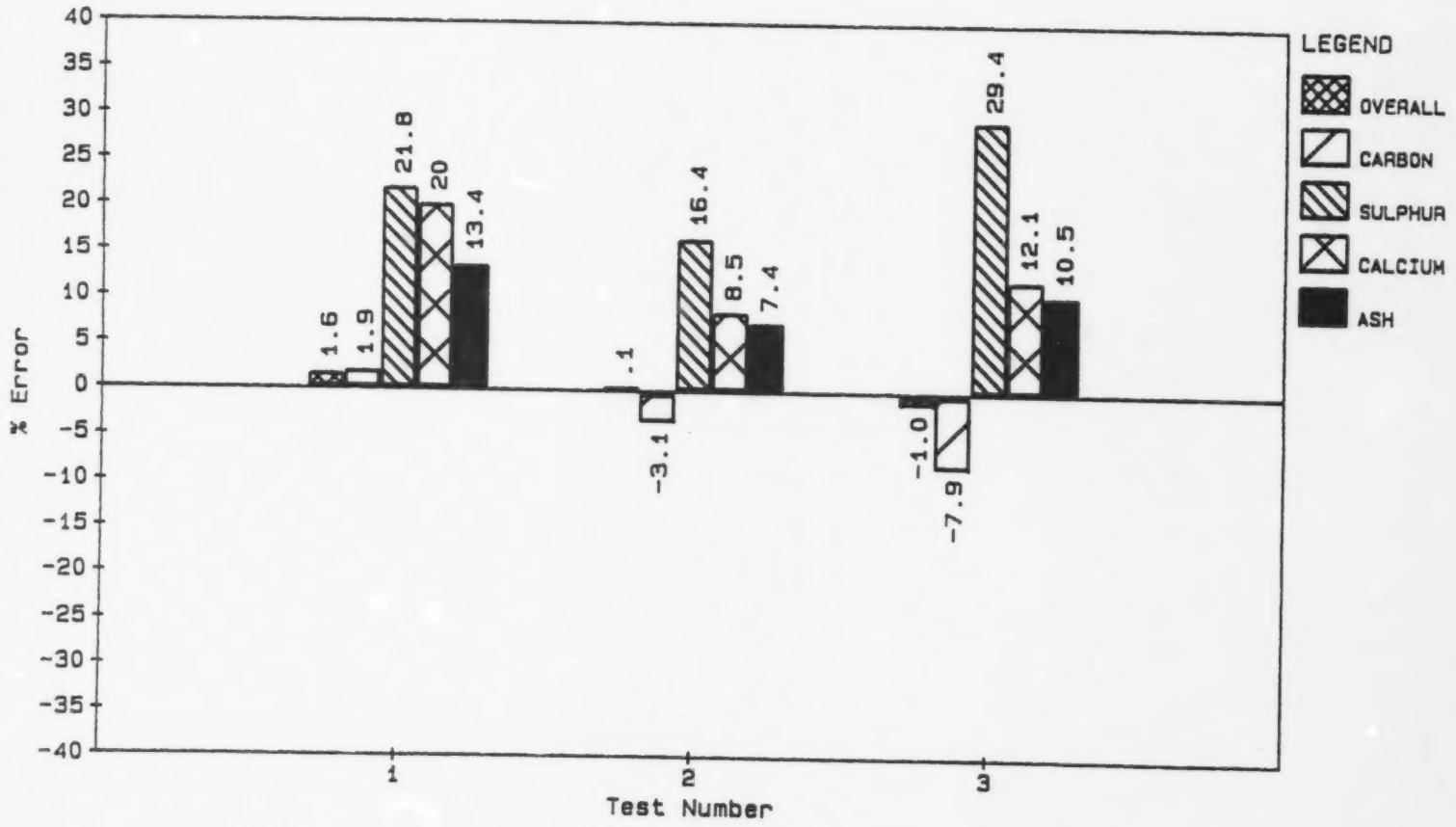


FIGURE 24

MASS BALANCE ERRORS

TASK II

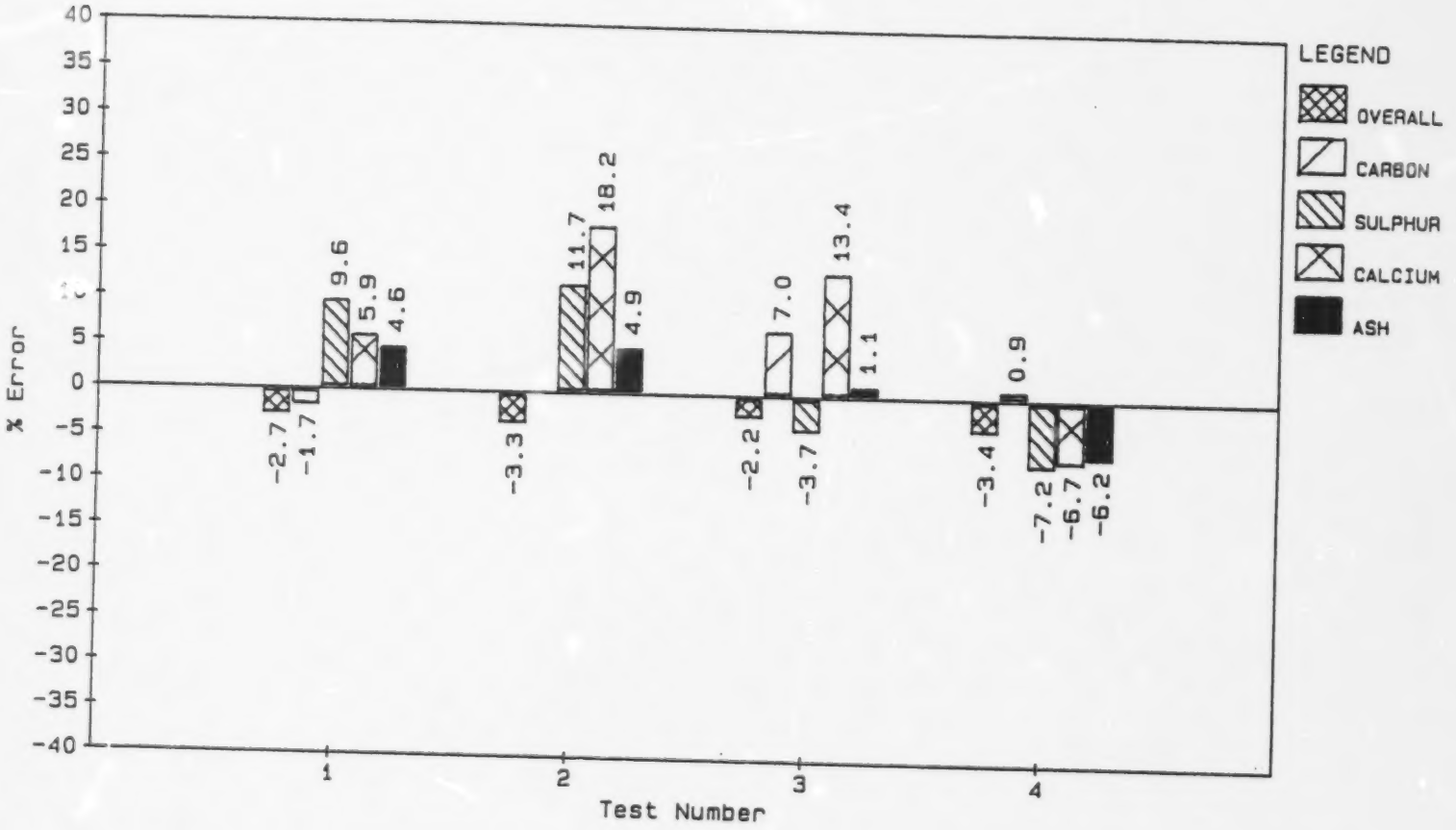


FIGURE 24 (cont'd)

MASS BALANCE ERRORS

TASK II

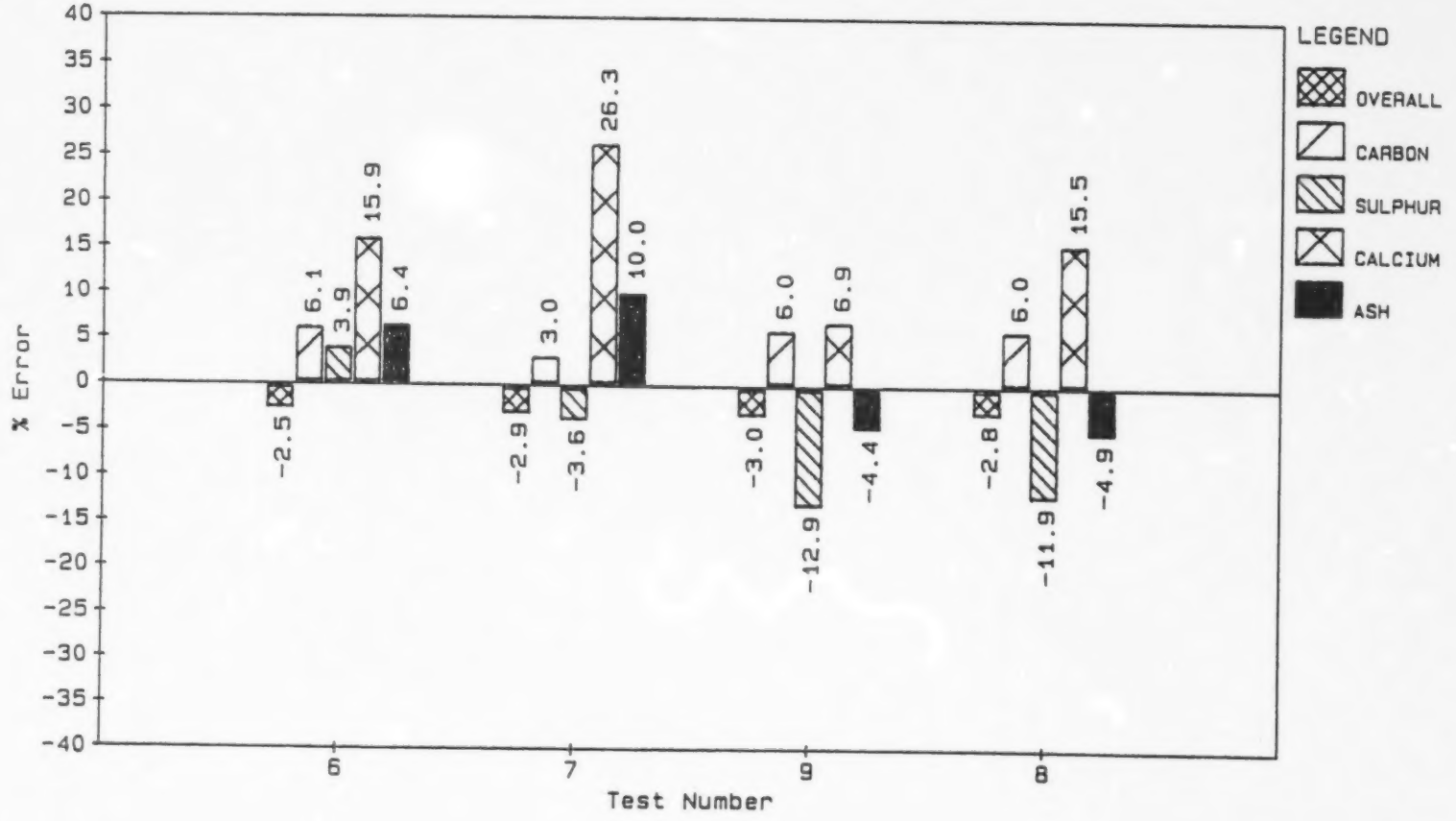


FIGURE 25

MASS BALANCE ERRORS

TASK III

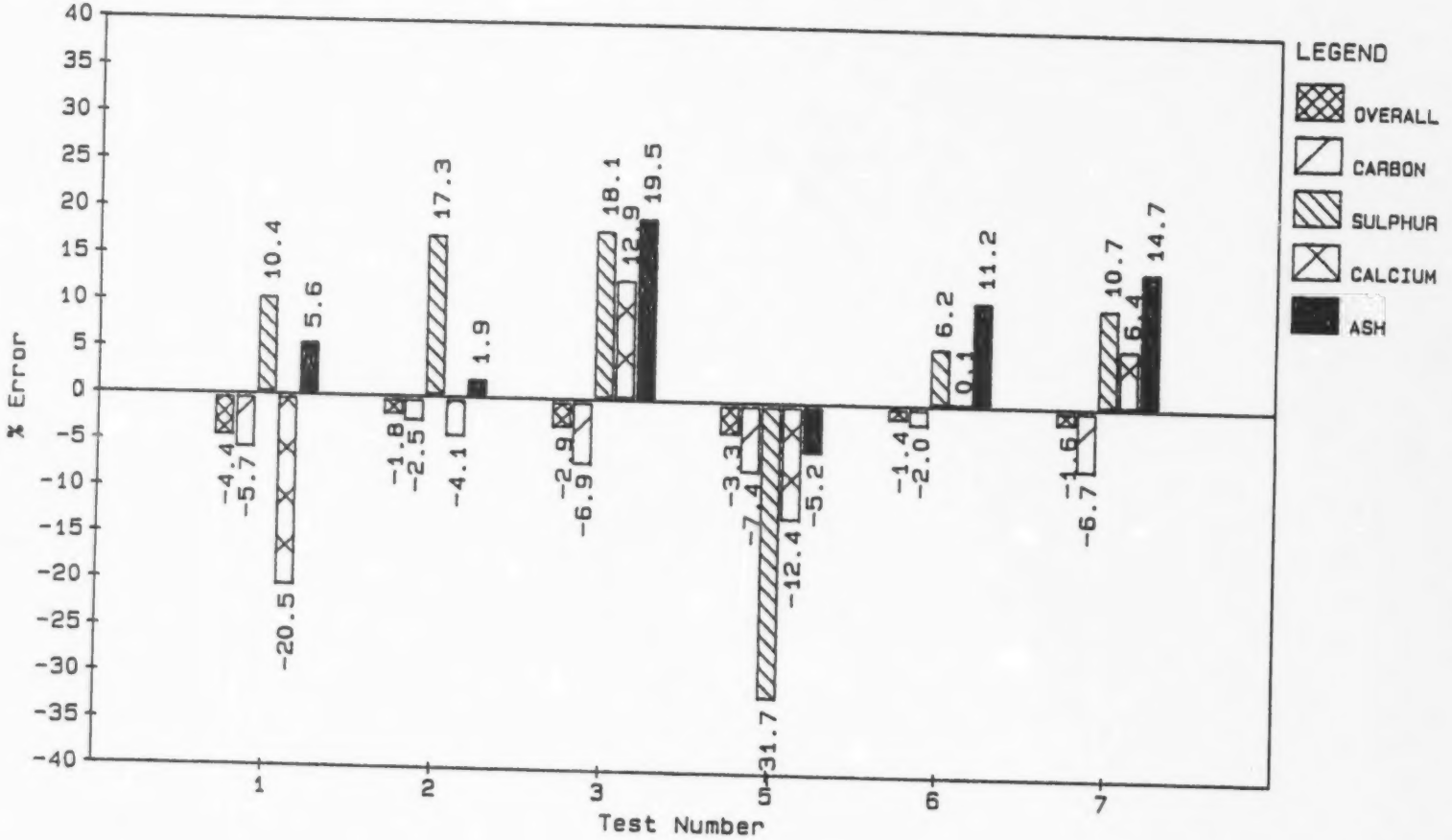
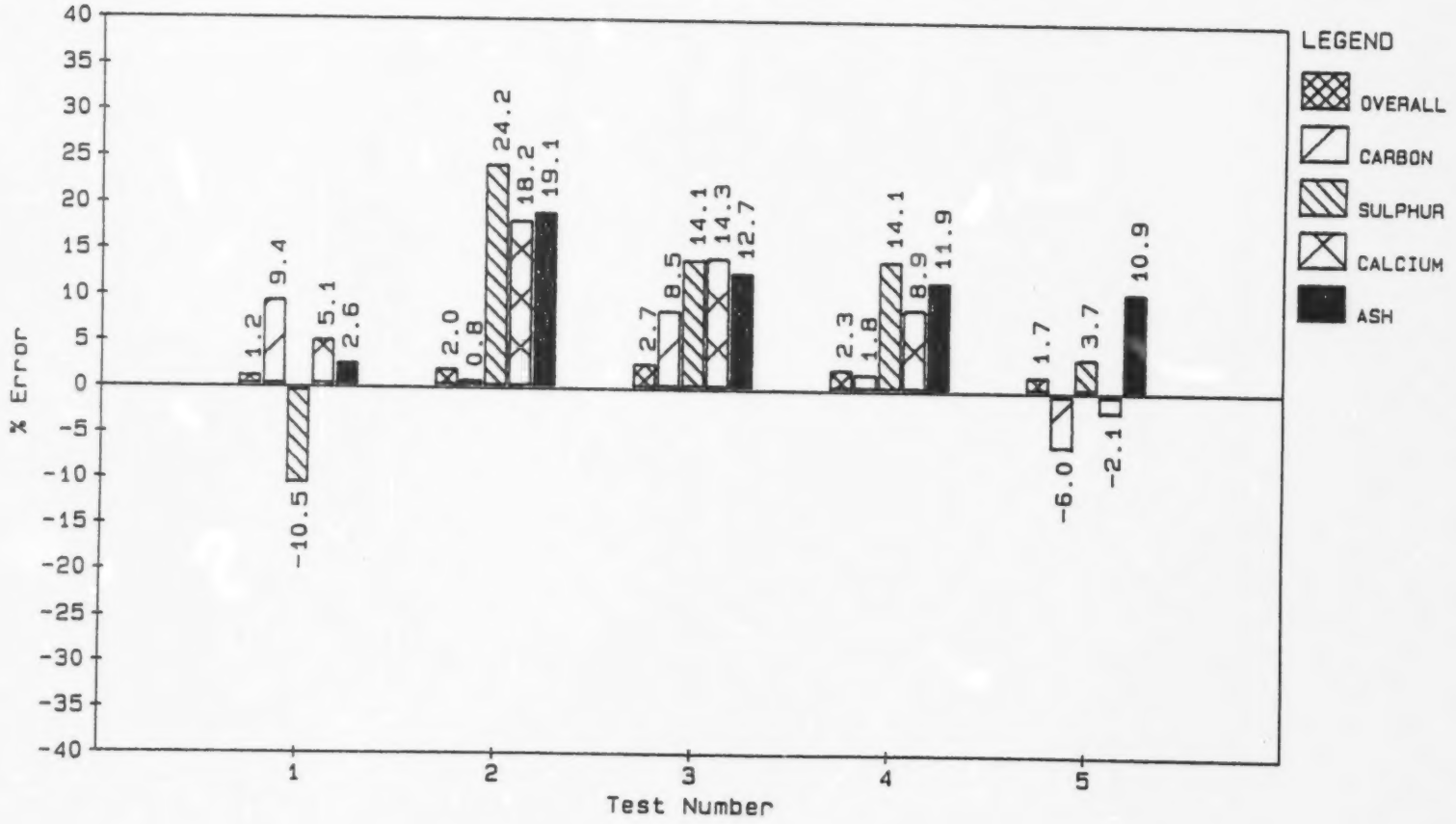


FIGURE 26

MASS BALANCE ERRORS

TASK IV



the test conditions. The statistical analysis of the baghouse discharge for Task V is shown in Table 10. Test 1 to 6 were carried out sequentially. In Test 1, the hourly discharge showed a wide variation between 0.1 and 11 and the total discharge was less than expected. The same pattern was repeated in Test 2, however the total discharge was more than expected. In Tests 3 to 5 the discharge was again less than expected, while in Test 6 the accumulated ash in the baghouse began to discharge at a very high rate. An example of an extreme case, the recorded baghouse discharge was 48.1 kgm in one hour. This is equivalent to approximately five times the hourly feed of coal and limestone.

The mass balance errors for Task V are compiled in Table 11 and are shown graphically in Figures 27 and 28. The actual results are given first and as can be seen, the mean values for the errors are quite acceptable, however, the standard deviation and the ash errors (see Figure 27) are extremely high. Also, for Test 6 there is far more ash leaving the system than fed into it.

The mass balance calculations were carried out once more but with the baghouse discharge arbitrarily adjusted to give an ash balance error of less than 5%. The mean of the errors shown in Table 11, are not significantly different from the actual baghouse results, however there was vast improvement in the standard deviation. Such improvement can be seen clearly by comparing Figures 27 and 28.

The problems of erratic baghouse discharge should be solved for future work. A hammer could be installed on the baghouse; the duration of the experimental runs could be extended or the bottom cone of the baghouse could be redesigned.

TABLE 10

STATISTICAL ANALYSIS OF BAGHOUSE DISCHARGE FOR TASK V

Test No.	Mean Value	Maximum	Minimum	Standard Deviation
1	1.58	11	0.1	2.08
2	3.24	8.9	1	2.19
3	1.4	2	0.3	0.36
4	1.69	3.3	0.4	0.87
5	1.35	2.7	0	0.8
6	6.61	48.1	1.5	11.36

TABLE 11

MASS BALANCE ERRORS FOR TASK V

Test. No.	Overall	Component Error, wt% Carbon	Sulphur	Calcium	Ash
Actual Baghouse					
1	-4.4	15.4	22.6	24.4	28.8
2	-6.9	-2.0	-28.0	-33.7	-6.6
3	-5.3	7.2	20.0	22.3	27.4
4	-4.7	4.9	24.1	22.6	20.4
5	-5.2	4.3	30.3	34.4	30.8
6	-10.9	-9.7	-82.9	-153.1	-83.3
Mean	-6.2	3.4	-2.3	-13.9	2.9
Standard Deviation	2.4	8.5	44.8	72.4	44.4
Corrected Baghouse					
1	-5.3	12.7	-0.3	-6.3	4.5
2	-6.3	0.5	-14.0	-16.7	4.8
3	-6.2	4.5	-3.4	-7.1	4.6
4	-5.3	2.9	8.9	3.3	4.5
5	-6.3	0.9	6.6	5.6	4.4
6	-6.5	5.5	8.1	-8.2	3.8
Mean	-6.0	4.5	1.0	-4.9	4.4
Standard Deviation	0.5	4.5	8.8	8.2	0.3

FIGURE 27

MASS BALANCE ERRORS

TASK V (Actual Baghouse)

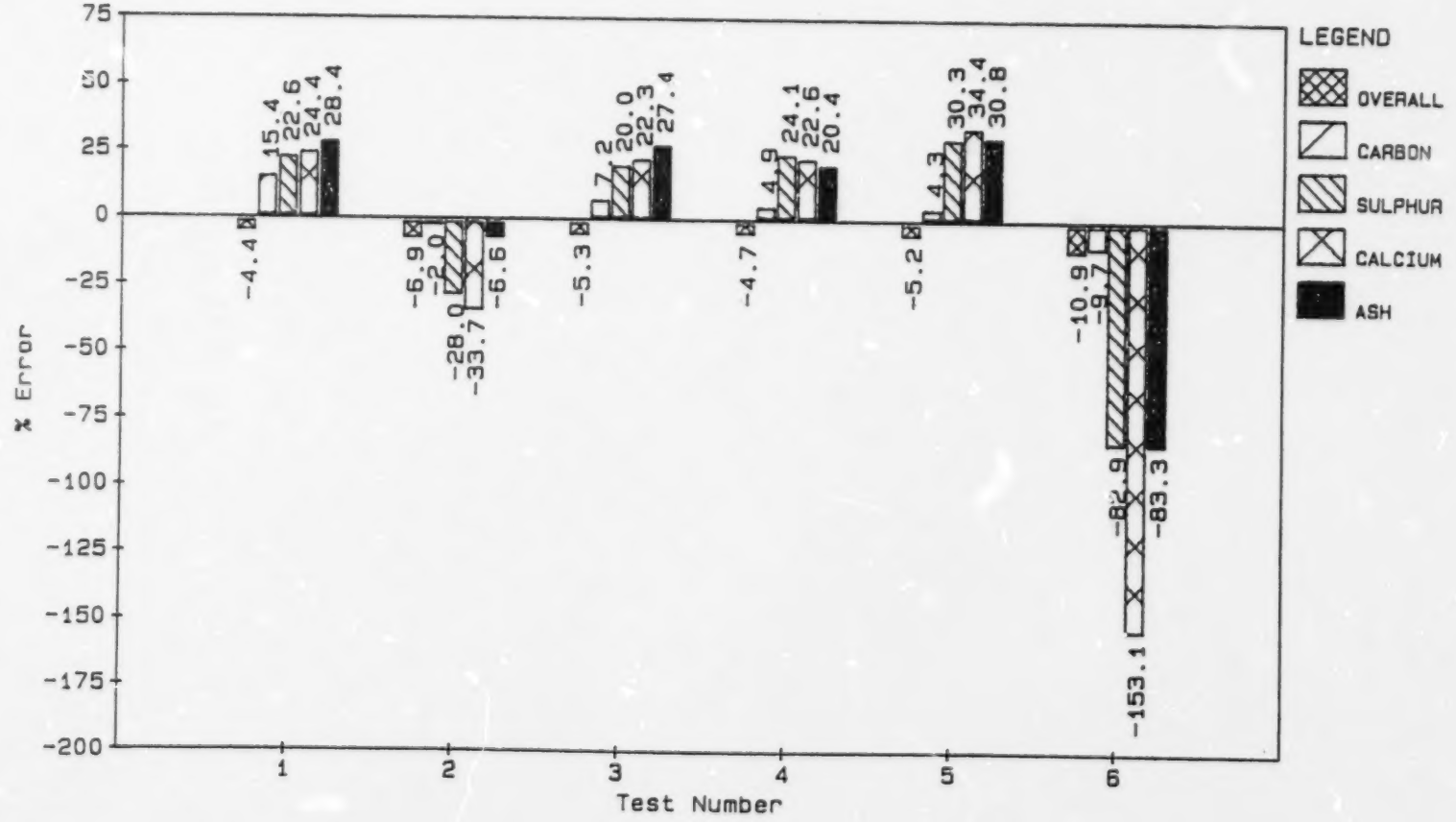
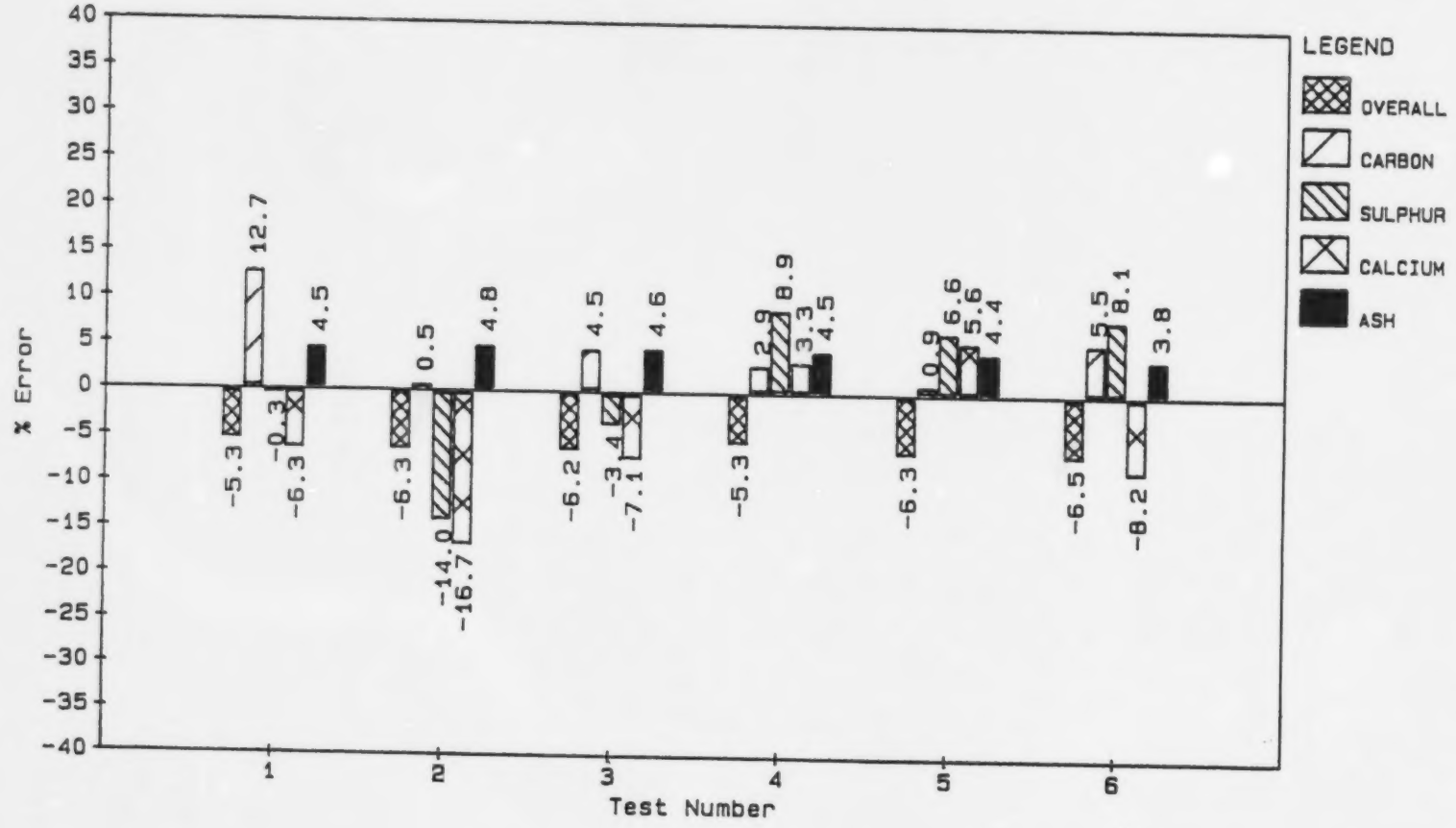


FIGURE 28

MASS BALANCE ERRORS

TASK V (Corrected Baghouse)



4.5 Dynamic Response of the Combustor

Stimulus response techniques are a powerful tool for investigating process dynamics. The particular stimulus used in this study was a step change in the limestone feedrate and the combustor response was taken as SO_2 concentration in the flue gases. As mentioned earlier in Section 4.1.1, the main objective was to gain some understanding of the relative importance of fresh and aged calcium particles.

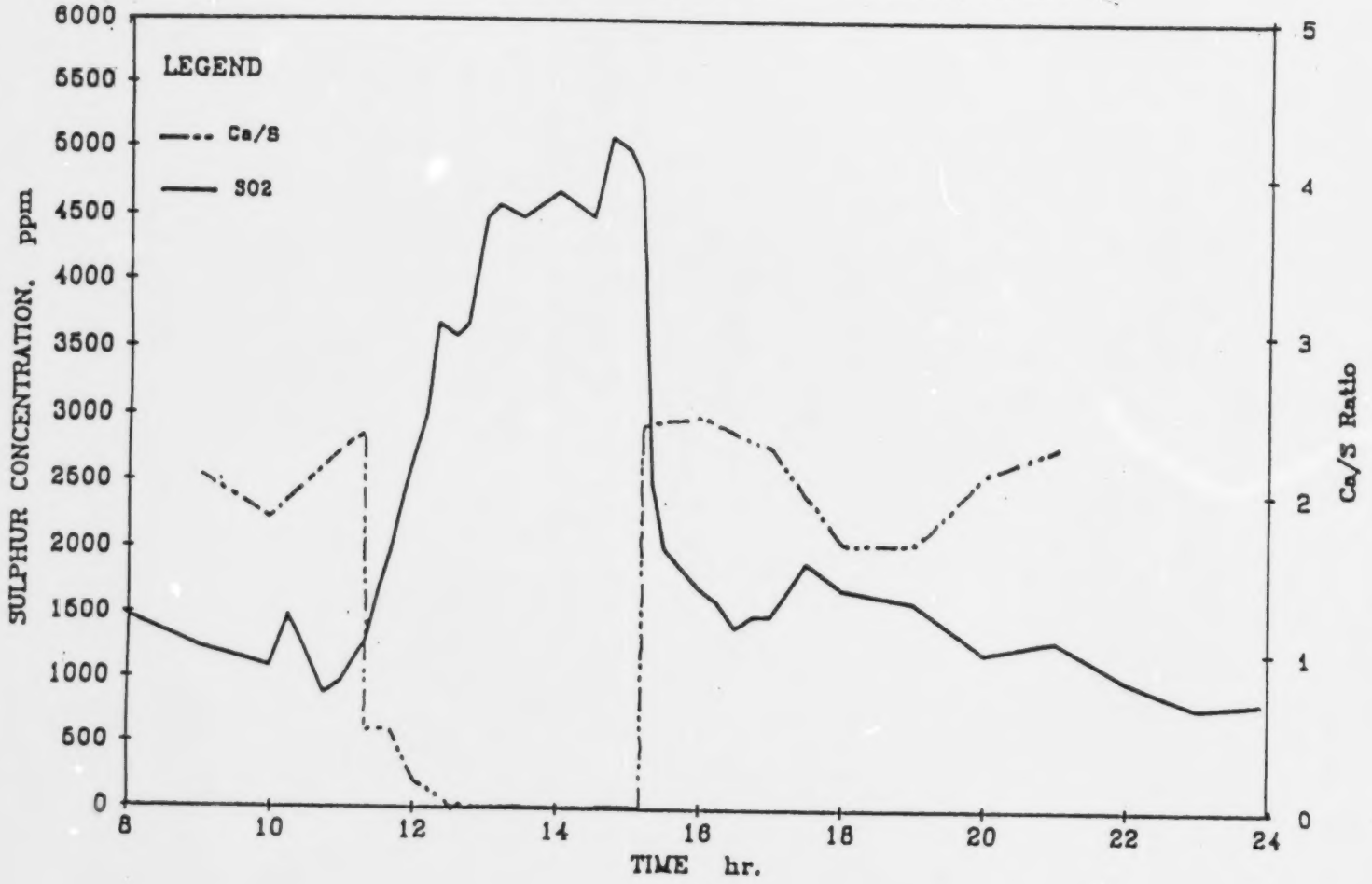
As this work was outside the scope of the present investigation, it was limited to a single experimental run. For this particular test the limestone was not blended with the coal but was supplied into the combustor using a separate feeder. The feedrate of coal and limestone was adjusted for a nominal Ca/S ratio of 2:1. After a baseline was established, limestone feed was stopped for approximately four hours, then it was re-initiated in order to re-establish the baseline conditions. The results for these three phases are given in Tables 1 and 2 as Tests 4a, 4b and 4c*.

The combustor responses are shown graphically in Figure 29. It shows that once the limestone feed was stopped there was an immediate rapid increase in SO_2 followed by a more gradual increase. In the first hour SO_2 concentration rose from approximately 1250 ppm to 3500 and in the following three hours the concentration increased to approximately 5000 ppm. The increase in SO_2 concentration in the first hour represented 60% of the total change in SO_2 concentration (1250 to 5000 ppm). Under the test conditions, 5000 ppm is approximately 90% of the theoretical SO_2 concentration at zero sulphur capture. It is important to note that four hours after stopping the limestone feed the system still contained an appreciable amount of calcium (see the chemical analysis for test 4b given in Volume II), however, for all practical purposes the calcium particles are uneffective in capturing sulphur capture.

* The data for Test 4 was divided into three sub-sets, each of which was considered as a steady state experiment. This may be valid for Test 4a, but it is not so for the other two phases. Unsteady state calculations would be necessary for such experiment, however, this could not be justified for a single experimental run.

FIGURE 29

Combustor Response to a Step Change in Limestone Feedrate



Once the feed of limestone was resumed the drop in SO_2 concentration was immediate and very sharp. The change in the first hour is approximately 88% of the final drop in SO_2 concentration which took approximately six hours to reach. After the initial drop, the gradual reduction in SO_2 concentration is due to accumulation of fresh calcium particles in the system.

The overall system behaviour suggests that the fresh calcium particles in the feed is far more effective than the aged particles. Also, beyond a certain age (4 to 5 hours for this particular test) calcium particles become very unreactive.

The above results are semi-quantitative at best. To gain more insight into the system dynamics, more sophisticated data acquisition and analytical tools will be required. Also, there are other stimulus signals other than the step function which are considered more powerful and more economical to use interims of the amount of information gained in the available experimental time (3 and 4).

4.6 Pressure Drop Measurements

One of the requirements in the statement of work (see Appendix A) was to instrument the combustor and the standpipe for differential pressure measurements for the extraction and plotting of pressure profiles. In consultation with the Scientific Authority, the system was fitted with 13 pressure gauges at the locations shown in Figure 4.

All the lines leading from the combustor to the gauges were purged with air in order to prevent line blockage. The air flow to each of the purge lines was adjusted before data collection began. For the combustor side, differential pressure gauges D1 to D9, the circulation of solids as well as the solids feed were stopped and the flow rate of air to the individual purge lines was adjusted such that all the differential pressure gauges were reading zero. For the stand pipe side (D10 to D13) the aeration and lift air were stopped and the purge air flows were adjusted for zero readings. Once this procedure was completed the gauges were ready to record the actual pressure drop due to solids loading.

A typical set of data is shown in Table 12. One of the main difficulties encountered in taking such measurements, was the wide fluctuation in pressure particularly for the standpipe signals (D10 to D13). In this form the recorded data is not useful because a pressure gauge is unsuited to this type of widely fluctuating signals. Another problem was the frequent blockage of the purge lines particularly those installed at the bottom half of the combustor. Following each blockage, the blocked line had to be blown clear and the gauges were re-calibrated (zeroed) as described above before pressure drop measurements could resume.

Considerable time and effort was invested in attempts designed to improve system reliability. Several designs of purge line tips were tested. These efforts did improve the system reliability somewhat, however, the main stumbling block was the limited amount of air available for purging and the fairly large number of lines to be purged. Total air flow is governed by the required operating velocity. This fixed amount of air was distributed, 60% as primary air, 10% as secondary air and 30% others. The latter was distributed among aeration, lift , pneumatic transportation and purges. The maximum available air

TABLE 12

Pressure drop * data recorded at hourly intervals using differential pressure gauges during Test 1 of Task 1

Gauge No.	Time, hour 1	2	3	4	5	6	7	8
1	1-2	0-2	1-2	1.5-3	1.5-2	1.5-2	1.5-2	1.5-2
2	2-3	1.5-5	1.5-2	2-2.5	2-2.5	>5	>5	>5
3	---	>5	>5	>5	>5	>5	>5	>5
4	---	>5	>5	>5	>5	>5	>5	>5
5	---	>5	>5	>5	>5	>5	>5	>5
6	---	---	---	---	---	---	---	---
7	3-7	0-10	1-6	0-10	4-10	2-8	2-8	2-7
8	0-6	0-8	0-10	0-6	0-10	2-8	0-10	2-7
9	0-3	0-4	0-4	0-3	0-3	0-3	2	3
10	>30	0-10	12	12	12	12	12	12
11	0-20	0-15	0-30	0-20	0-30	0-35	0-10	0-25
12	10-70	0-60	0-50	10-100	0-75	10-50	0-30	0-30
13	10-100	10-100	0-50	30-130	0-75	10-30	0-30	0-30

* Inches H_2O

flow for purge was approximately 5% of the total air flow. This problem can be resolved by reducing the number of gauges, changing the operating conditions (air distribution) or on scale up.

As a final attempt some of the pressure gauges were replaced by pressure transducers and the output was plotted on a chart recorder. This measure proved successful and useful information was gained on the system behaviour particularly on the operation of the J-valve.

The two traces shown in Figure 30 are for Transducers D13 and D12 (see Figure 4). The traces show clearly that the J-valve work on pulses which are approximately 1 minute apart. Each pulse would transfer a slug (a batch) of material into the combustor. A possible mechanism is that a small air pocket (bubble) would form within the solid phase. The bubble would grow larger and larger until it finally collapses, creating a disturbance and displacing the solids across the horizontal section of the J-valve and toward the lift air port. The lift air would carry the solids upward and into the combustor. This cycle is then repeated at approximately 1 minute intervals.

Figure 31 shows the traces for pressure Transducers D2 and D3, located near the top of the combustor (see Figure 4). These two traces show clear evidence of a periodic surge in pressure indicating increase in the solid loading. The pressure peaks are at approximately 1 minute intervals and it is certain that these are caused by the pulsing (discontinuous) transfer of solids from the standpipe into the combustor.

Figure 32 shows traces for transducer 8 and 9 located at the base of the combustor. These traces show the same periodicity of the previous traces, however, the signal is much noisier indicating a denser solid phase. This result is consistent with typical solid loading (pressure drop) profiles showing a fairly dense phase at the base of the combustor and a low solid loading at and near the top.

For more rigorous analysis, the data should be acquired, stored and digitized. In which case, well known signal process techniques can be applied. For example cross-correlating signals from transducers 13 and 2 would give the time lag between the two signals which is the time required for a slug to travel through the combustor. This is

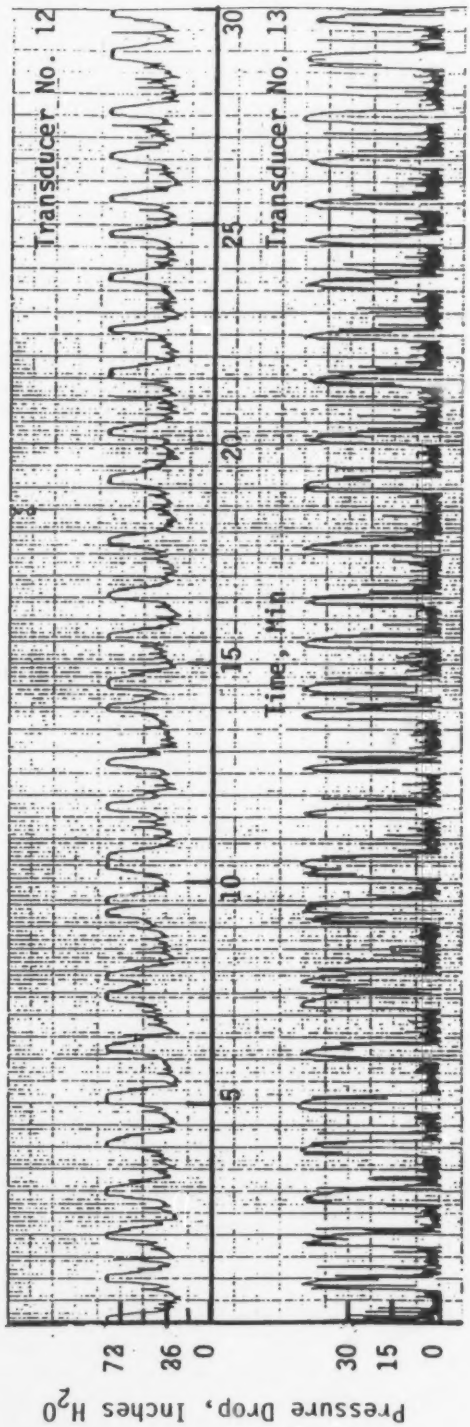


FIGURE 30: Pressure drop record showing the output of transducers 12 and 13 located at the base of the J-valve.

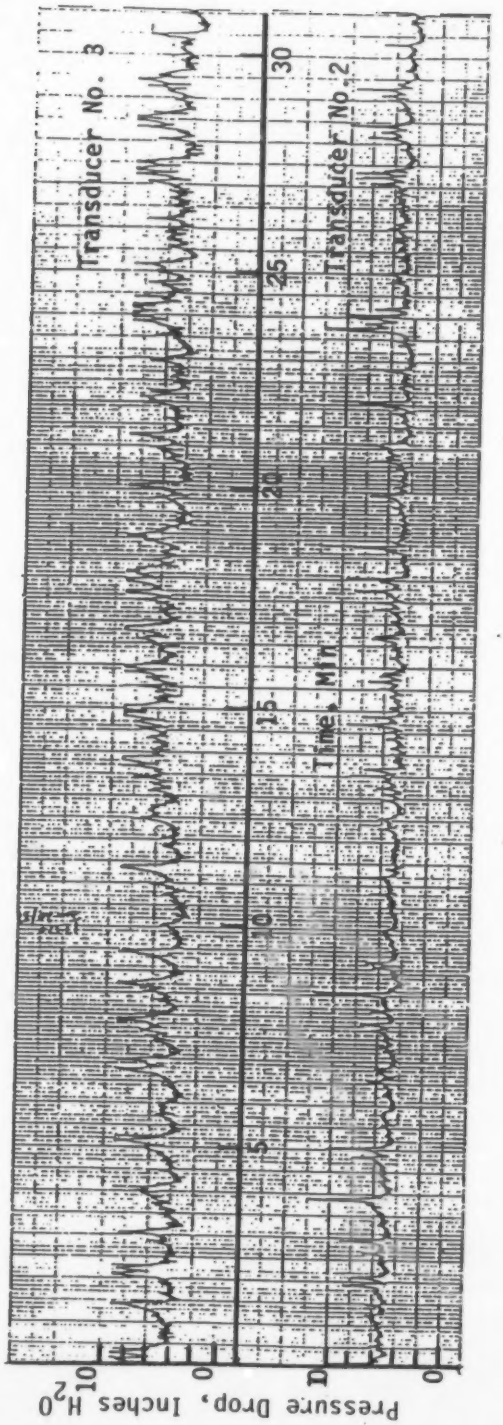


FIGURE 31: Pressure drop record showing the output of Transducers 2 and 3 located near the top of the combustor.

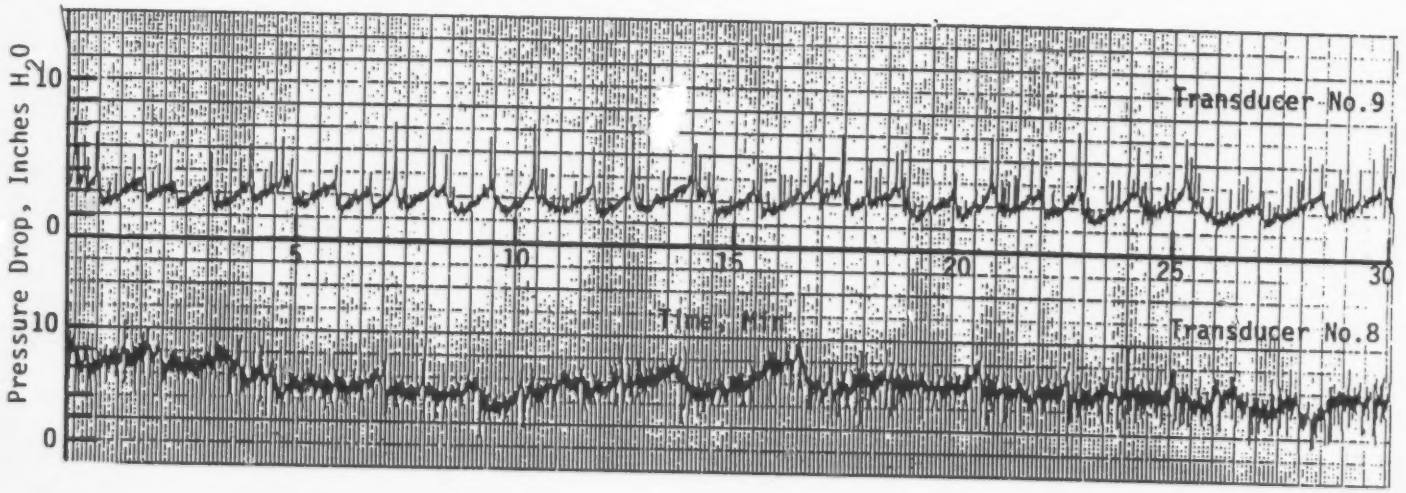


FIGURE 32: Pressure drop record showing the output of Transducers 8 and 9 located at the bottom of the combustor.

equivalent to the residence time of solids in the combustor. Autocorrelation of any trace would give the period at which the signal repeat itself. Others which can be applied include filtering, averaging and frequency analysis (5).

5.0 CONCLUSIONS AND RECOMMENDATIONS

In this section the general conclusions which have emerged from this study will be given, as well as recommendations for future work.

5.1 Conclusions

The main conclusions of this study are as follows:

1. Calcium utilization and sulphur capture can be improved significantly by screening off relatively large and very fine particles. The most effective size range for this particular system appears to be between 600 to 40 μm.

2. Based on current results, there is no significant differences between the reactivity of Havelock and Elmtree limestones.

3. Burning Minto coal with Havelock limestone at a Ca/S ratio of 2:1 could result in a sulphur capture in excess of 85%.

4. The best sulphur capture was achieved at a bed temperature of 850°C. sulphur capture declined marginally at 900°C and significantly at 800°C.

5. Increasing the operating velocity from 7 to 7.9 m/s led to approximately 10% decrease in sulphur capture.

6. Recycling of the dust from the secondary cyclone at a high rate (5 times the feed rate) led to a significant increase in sulphur capture (from approximately 73 to 87%).

7. Sulphur capture is not sensitive to recirculation rate in the range of 600-1300 kg/h. Much higher recirculation rate may lead to a decline in SO_2 capture and combustion efficiency.

8. Aged calcium oxide particles are less reactive than fresh ones. Based on the limited data available and under the conditions tested, calcium oxide particles become very unreactive after 4-5 hours.

9. For this particular J-valve design, the solid recirculation from the standpipe into the combustor is not in a continuous flow, but it does occur in pulses. The solid loading across the entire height of the combustor is affected by these pulses. The pressure drop (an indication of solid loading) at all locations in the combustor appears to peak, then decay at regular intervals consistent with the frequency of the J-valve pulses.

10. Within the range tested a variation of oil shale/coal ratio and shale particle size did not have a significant effect on sulphur capture. The effect of these parameters, however, may have been masked by the presence of limestone in large concentration.

11. At the same Ca/S ratio there appears to be no significant difference between the use of oil shale or limestones. However, the Oil Shale used had a low calcium content and was supplemented with a significant amount of limestone.

12. For Nova Scotia feedstocks, Windsor limestone is more reactive than Glencoe. For both limestones the -35 mesh size fraction gave marginally higher sulphur capture than the -18 mesh size fraction.

5.2 Recommendations for Future Work

Future research and development work is required in order to enhance our current understanding of number of important issues. Those identified in this study are as follows:

1) Limestone particle size is one of the most critical parameters affecting sulphur capture. As discussed in Section 4.1.1, particle size optimization is a complex process. A number of approaches can be used, this may include thermogravimetric

analyses, pilot scale and demonstration scale CFBC studies, each of which has its own advantages and limitations. A pilot scale test program may involve testing of a number of narrow size fractions of a particular limestone under a given set of operating conditions. A number of limestones should be tested in order to quantify the impact of limestone reactivity and hardness (attrition resistance) on the optimum particle distribution.

2) Recirculation of the ash from the secondary cyclone into the combustor have been shown to increase sulphur capture significantly. The rate of ash recirculation should be optimized in order to avoid excessive unproductive dust recycle. Also, the active constituents in the recycled material should be identified. This could be the fly ash itself, which can react with the sorbent and increases its reactivity, or it could be a highly reactive CaO particles.

3) Methods for increasing the reactivity of natural sorbents and fly ash should be investigated. Additives such as chlorides of Na and K are known to bring about distinct improvements in sorbent reactivity. Pretreatment of sorbents by the addition of aquease $NaHCO_3$ or the hydration of fly ash by steaming prior to injection into the combustor are options which should be explored. Fly ash from different sources including the Chatham Demonstration unit should be investigated.

4) Oil shale of varying and higher calcium content than the particular batch used in this study (2.86% calcium) should be tested. This would provide the necessary results required for the evaluation of effectiveness of oil shales in controlling sulphur emissions from CFBC units burning high coals.

5) Tracer stimulus response techniques have been previously applied to the study of chemical reactor design, residence time distribution in combustors, mixing, turbulence, etc. These techniques are powerful tools and can be applied to CFBC units for better understanding of the process dynamic and transfer functions which are indispensable for process optimization and controllers design.

6) Detailed pressure drop measurement should be carried out using pressure transducers coupled with data acquisition and storage system. Using the appropriate signal processing techniques could lead to better understanding of solid loading profiles, solids residence time distributor and possibly the identification of the degree of mixing at different regions of the combustor.

7) Correlation of the results of this study, with those from the demonstration scale unit at Chatham, is perhaps the most important next task. Once this has been accomplished, the RPC unit with its fast turn around time, flexibility and cost advantages can be used with confidence for fuel and sorbent characterization and process optimization.

6.0 ACKNOWLEDGEMENTS

The project was jointly funded by the Federal and Provincial Governments, through EMR Cost Shared Program and the Industrial Innovation and Technology Development Subsidiary Agreement.

Special thanks are due to Don Gemmell, Department of Commerce and Technology, Frank Friedrich and Ivan T. Lau, CANMET, and Rod Pickles, N.B.E.P.C. for their support of the project. The assistance of Ross Gilders, Lech Lewczuk and other technical personnel particularly Camille Doucet in carrying of the work is gratefully acknowledged.

7.0 REFERENCES

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2. **Badin, E.J. and Frazier, G.C.**, "Sorbents for Fluidized-Bed Combustion",Environment Sci. Tech., Vol. 19, No. 10, pp 894-901, 1985.

3. **Turan, A., Swithenbank, J., Billings, S.A., and Salib, P.**, "Pseudo-Random Stimulus Response of Combustion Systems", Journal of Institute of Energy, p 181-190, Dec. 1980.

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5. **Beauchamp, K.G.**, "Signal Processing Using Analog and Digital Techniques", George Allen and Unwin Limited, London, 1973.

8.0 AUTHORIZATION

AUTHORIZATION

DEPARTMENT: MINERALS AND FUEL TECHNOLOGY

PROJECT TITLE: "Investigation of Sulphur Capture During Circulating Fluidized Bed Combustion of Eastern Canadian Coals"

PROJECT NUMBER: 08008016/0001

PROPOSAL NUMBER AS/87/30P

REPORT MDP/88/17 (revised)

Prepared By: _____
P. Salib

Reviewed By: _____
D. Abbott

Approved By: _____
Dr. R.S. Boorman

Date: March 28, 1989

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research and productivity council
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March 28, 1989

Mr. Don Gemmell
Dept. Commerce & Technology
P.O. Box 6000
Fredericton, NB
E3B 5H1

Dear Mr. Gemmell:

RE: Investigation of Sulphur Capture During Circulating Fluidized Bed Combustion of Eastern Canadian Coals

Please find enclosed 5 copies of the final report on the above mentioned contract.

This report is a revision of the earlier draft, taking into account your comments and suggestions.

I would like to take this opportunity to thank you for your help and co-operation and I look forward to working with you on future projects.

Yours sincerely,

P. Salib, P. Eng.
Head, Fuel Technology
Minerals & Fuel Technology

/cem

Enclosures

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research and productivity council
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March 28, 1989

Mr. Ivan T. Lau
CANMET - ERL Division
Dept. of Energy, Mines & Resources
555 Booth Street
Ottawa, ON
K1A 0G1

Dear Mr. Lau:

RE: Investigation of Sulphur Capture During Circulating Fluidized Bed Combustion of Eastern Canadian Coals

Please find enclosed 5 copies of the final report on the above mentioned contract.

This report is a revision of the earlier draft, taking into account your comments and suggestions.

I would like to take this opportunity to thank you for your help and co-operation and I look forward to working with you on future projects.

Yours sincerely,

P. Salib, P. Eng.
Head, Fuel Technology
Minerals & Fuel Technology

/cem

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APPENDIX A
WORK STATEMENT

WORK STATEMENT

RPC Proposal No. As/87/30P, dated April 10, 1987 is considered an integral part of the work statement and it is included in this Appendix. The basis of payment for the contract, however, was different from that given in the proposal. A summary of costs are given in Table A-1 and it replaces Table 6 on page 27 of the proposal.

The test matrices for all five tasks were formulated in consultation with both Commerce and Technology and CANMET representatives. There was no deviation from the statement of work, except perhaps for Task I which was modified in full consultation with the above mentioned representatives.

STATEMENT OF WORK

TITLE: INVESTIGATION OF SULPHUR CAPTURE DURING CIRCULATING FLUIDIZED BED COMBUSTION OF EASTERN CANADIAN COALS [Part of Energy, Mines and Resources 50/50 Cost Shared Program and part of the Program of the Federal Panel on Energy Research and Development (PERD)]

TASK 1

(1) Write up and acquire approval of the sampling and analysis procedure with C & T and Canmet representatives:

(2) Install pressure probes on the combustor and return leg to facilitate the extraction and plotting of pressure profiles.

(3) Using air factors of
Primary: normally 60%
Secondary: normally 10%
Others: normally 30%

(4) Maintaining temperature range
845-875° C

(5) Ca/S molar ratio of 2:1

(6) Utilizing Havelock limestone

(7) Utilizing New Brunswick coal

(8) Utilizing recycle rates as per phase 1, make measurement of the same. Carryout 2' runs with the rig as configured and three additional runs bypassing the secondary cyclone. Upon completion verbally discuss results with Commerce and Technology representative who will liase with CANMET and N.B.Power.

TASK II: As proposed Task 2.1 to 2.5, except that dependant on sulphure capture acheived in Task I without the secondary cyclone, tests runs will be attempted at higher ratios if 80% minus are achieved and lower if 95% are acheived but not both.

TASK III: Utilizing the optimization of both fuel and absorbant, and molar ratio from Task II investigate bed temperature variation, gas velocity and recycle rates, and its affects on combustion efficiencies and sulphure capture

TASK IV: Includes Task I to III inclusive only utilizing New Brunswick coal with oil shale. This requires prior consultation with C & T and N.B. Power to set the parameters which N.B. Power will have from the Chatham plant by the time of implementation of this task.

Task V: Analysis and report writing.

Additional Contact conditions as follows:

(1) Tests are to be run as outlined and any changes require prior written approval of C & T and Canmet Officials.

(2) Non-compliance with item one will result in non payment for work carried out.

TABLE A-1
SULPHUR CAPTURE INVESTIGATION PHASE II
SUMMARY OF COSTS

1. **Direct Labour Charges**

Category	Rate	Hours	Cost $
P5	84	90	7,560.00
P4	79	1200	94,800.00
T2	42	800	33,600.00
T1	27	1837	49,599.00
LT1	27	1113	30,051.00*
TOTAL DIRECT LABOUR			215,610.00

2. **Direct Charges**

Material and Supplies	8,493.50
Drafting	500.00
Travel	1,000.00
TOTAL DIRECT CHARGES	9,993.50
TOTAL COSTS	$225,603.50

* Refers to total costs of chemical analysis which are determined on a per determination basis. This cost has been converted to a man-hour figure based on an LT1 rate.

DSS File No. 06SQ-023440-6-9060X

INVESTIGATION OF SULPHUR CAPTURE DURING CIRCULATING FLUIDIZED BED COMBUSTION OF EASTERN CANADIAN COALS

A Proposal Submitted

to

Supply and Services Canada
for funding under the
CANMET ENERGY CONVERSION
COST SHARED PROGRAM
1987-88

and

Science & Technology Secretariat
Government of New Brunswick

PROPOSAL NO.: AS/87/30P
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1.0 SUMMARY

The combustion of various coals with limestones has been demonstrated in the circulating fluidized bed combustor (CFBC) in the pilot plant of the New Brunswick Research and Productivity Council. The sulphur capture obtained during these runs ranged from 50 to 97 percent. These tests indicate the need for additional work to identify and optimize the operating parameters affecting sulphur capture in a CFBC.

This proposal is a revised version of RPC's earlier submission (Optimization of Sulphur Capture During Circulating Fluidized Bed Combustion of Minto Coal, RPC Proposal No. AS/87/15P dated November 10, 1986). The revisions follow discussions with D. Gemmell, Province of N.B. and F. Friedrich CANMET on the correlation of the performance of the RPC unit with that of the Chatham demonstration CFBC of NBEPC.

The project consists of two phases, the first being a modification and tests of RPC existing equipment, while the second would be a series of combustion tests on Minto and Prince coals, with Havelock, Elmtree and other limestones and N.B. oil shale. This work will demonstrate the full capacity of CFBC for sulphur capture during the combustion of high sulphur eastern Canadian coals. The results will be useful for furthering the combustion optimization of large scale CFBC units such as the Chatham facility.

Funding for this project is requested under the CANMET Energy Conversion Cost Shared Program 1987/88. The total estimated cost of the project is $249,674.00 The requested federal support is $112,801.75, which represents 50% of Phase II costs. The requested funding from the N.B. provincial government is $136,872.25. This would cover 100% of Phase I costs including $6,000 contingency and 50% of Phase II costs. The project can be completed 11 months after commencement.

2.0 TECHNICAL PROPOSAL

2.1 Introduction

The bituminous thermal coals of eastern Canada are high in sulphur. Those used in Nova Scotia generally contain more than 4 per cent sulphur and in New Brunswick, the Minto coal can have as much as 8 per cent sulphur. Growing sensitivities over air quality means that there is concern for the amount of sulphur dioxide discharged to the atmosphere and various control technologies are available. The bubbling fluidized bed is one which is being overtaken by circulating fluidized bed combustion (CFBC). The advantages of the latter include:

1. Improved carbon and calcium utilization. This means good combustion efficiency and less waste product for disposal.

2. Elimination of in-bed tubes which reduces the chances of tube failure problems.

3. Low NO_x emission.

The major investment by the federal government in a demonstration circulating fluidized bed unit at the Chatham NB plant of the New Brunswick Electric Power Commission reflects Canadian interest in using this technology.

RPC has led development in dealing with the high sulphur coals of eastern Canada. It showed the possibilities for sulphur control, energy generation and liquid petroleum production in the co-utilization of oil shales and eastern Canadian coals. RPC has constructed and commissioned, with federal and provincial financial assistance, a $0.5 million pilot plant containing a circulating fluidized bed combustor which can be used for fuel and combustion process optimization.

Since commissioning of the RPC circulating fluidized bed combustion pilot plant at the end of 1985 a number of coals have been burnt with different limestones.* Bituminous coals from Devco (NS), Minto (NB), and from the US (Pennsylvania) were burnt as well as a lignite from Poplar River (Sask.) Havelock and Elmtree limestones from New Brunswick were used as well as Exshaw limestone from Alberta. Calcium/sulphur ratios ranged from 1.5 to 2.5 and sulphur capture from 50 to 80 per cent. The relatively low sulphur capture was due to the fact that a fresh bed of sand had been used for individual runs which were limited to a maximum of 12 hours duration. The large amount of bed material contained in the standpipe of the CFBC at RPC, (approximately 250 kg) and the short test duration did not allow the standpipe bed to reach a steady state composition. An additional test which lasted for 5 days, was performed with a mixture of sand and limestone as the starting bed. The purpose of seeding the bed with limestone was to simulate a steady state CaO concentration in the bed. The sulphur capture during this test using Devco coal and Elmtree limestone was greater than 95 percent at a molar Ca/S ratio of 2.5:1.

The results from these tests showed that the recycling of bed material high in CaO concentration significantly improves the sulfur capture in the combustor. In a circulating fluidized bed the recycled material (from the standpipe) is large compared to the feedrate of fresh coal and limestone. Therefore recycling a standpipe bed containing unutilized CaO would greatly increase the actual Ca/S ratio in the combustor. For example, the expected concentration of unutilized CaO in the standpipe bed would be approximately 24% (see Appendix A) using Devco coal and Elmtree limestone with a molar Ca/S ratio of 2.5. Assuming a recycle rate of 2000 kg/hr in the CFBC, the actual Ca/S ratio in the combustor

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* Production of Circulating Fluidized Bed Combustion Wastes for Characterization" RPC report AS/87/03 dated April 24, 1986. Confidential report to Dearborn Environmental Consulting Services.

would be 811 (Appendix A). With such a large Ca/S ratio the sulfur reduction would be expected to be high. This is supported by the test performed at RPC with a bed seeded with limestone.

Other parameters which may have a significant effect on sulphur capture include sorbent type and particle size, Ca/S ratio, bed temperature, operating gas velocity, air distribution between primary and secondary streams, excess air levels and solid recycle rate. In this study some of these parameters will be examined. The main emphasis however, will be on characterization of the combustor performance in terms of SO_2 capture and combustion efficiency while operating on a number of coal and sorbent combinations.

Discussions have been held with D. Gemmell, Commerce and Technology, Province of New Brunswick, F. Friedrich and B. Anthony, Energy Research Laboratories CANMET, Ottawa about the performance of the RPC pilot plant CFBC and the role that it can play in lead-up work for the Chatham demonstration unit for the New Brunswick Electric Power Commission.

The work program suggested here is a revision of the test program on sulphur capture submitted in November 1986 for funding under the cost shared energy conversion program of the Federal government.

2.2 Method of Approach

2.2.1 Objectives

The main objectives of the study are as follows:

(i) Demonstrate and correlate the performance of RPC's pilot-scale circulating fluidized bed combustor with that of N.B. Power's demonstration unit at Chatham.

(ii) Assess the effect of limestone type and particle size on sulphur capture for selected coal/limestone combinations.

(iii) Assess the effect of air distribution on the combustor performance in terms of SO_2 capture, combustion efficiency and NO_x formation.

(iv) Assess the effect of oil shale particle size on sulphur capture.

2.2.2 Scope of the Work

The project will be divided into two phases. The first consists of modification and testing of RPC's combustor, while the second will be the actual test work necessary to meet the above mentioned objectives.

The need for combustor modification became apparent as a result of previous test work carried out at RPC, a summary of which is given in Appendix B. The modifications are necessary in order to minimize CaS formation by reducing the size of the standpipe (bed inventory) and redesign of secondary air ports. This would also reduce the time requirement to reach steady state operation. Detailed description of the work to be carried out in each phase is given in Section 2.3

2.2.3 Existing Test Facilities

The Research and Productivity Council's circulating fluidized bed pilot-scale test facility is shown schematically in Figure 1. The main components of the system are the combustor, a primary cyclone and a standpipe for solids recirculation, a cooling section, a secondary cyclone and a baghouse.

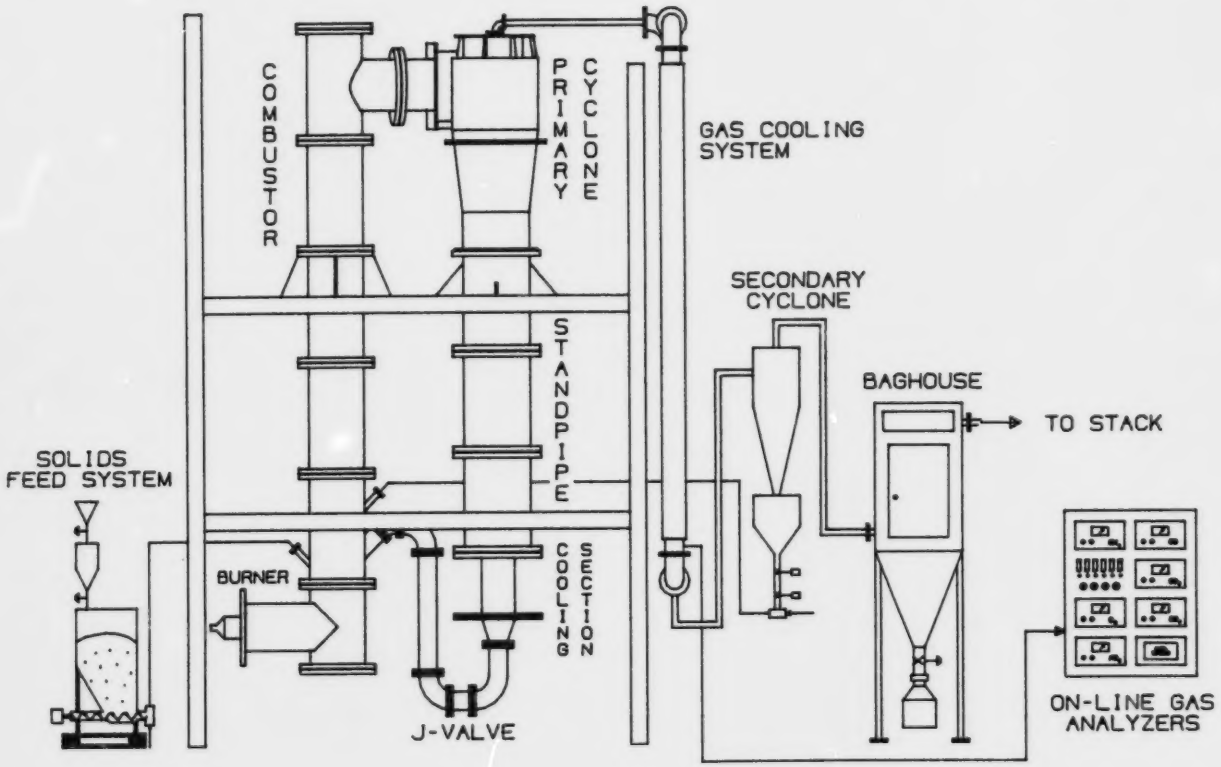

FIGURE 1: RPC CIRCULATING FLUIDIZED BED COMBUSTOR

The Combustor

The cylindrical combustor has a total height of 5.8 m (19'), and inside and outside diameters of 12.7 cm (5") and 50.8 cm (20") respectively. The wall has an inner layer of high temperature refractory , a middle layer of castable insulating refractory, and a carbon steel shell. A 0.5. MBtu propane burner is used for start-up purposes.

Primary air is fed into the bottom of the combustor via a distributor plate; a perforated plate with 10% open area located 66 cm (26") from the base of the combustor column. Secondary air is fed into the combustor via four diagonal ports at 205.7 cm (81") above the distributor plate.

Stand Pipe

The stand pipe is used for storage and recirculation of solids received from the primary cyclone. It is 2.7 m (9') high and has inside and outside diameters of 22.8 cm (9") and 61 cm (24") respectively. The stand pipe is also refractory lined and is also equipped with discharge and sampling ports.

A cooling section is installed at the bottom of the stand pipe. This is used to extract excess heat generated in the combustor so that both bed temperature and air/fuel ratio can be set and maintained at the appropriate levels.

Cyclones

The primary cyclone is 30 cm (12") in diameter, and is refractory lined in a similar manner to the reactor and the standpipe. The secondary cyclone is made of stainless steel and is 25 cm (10") in diameter. The solids collected by the secondary cyclone can be recycled pneumatically to the combustor, 48.3 cm (19") above the distributor plate.

The J-Valve

Circulation of solids from the standpipe to the reactor is achieved by the non-mechanical J-valve. Air is supplied at two points - one at the base of the standpipe, and the other at the base of the vertical section of the J-valve. The J-valve is made of 2", Schedule 40 stainless steel pipe. The solids from the J-valve enter the combustor at a location 36.8 cm (14.5") above the distributor plate.

Gas Cooling System

Off-gases from the primary cyclone are fed into a water cooled one pass heat exchanger. It is made of two concentric pipes of a nominal diameter of 15 cm (6") (Schedule 40) and 20 cm (8") (Schedule 60), respectively. The height of the exchanger is 5.2 m (17").

The Baghouse

The flue gases from the cooling section are fed through the secondary cyclone into the baghouse. The latter is a continuous duty self cleaning dust collector which consists of 16 high temperature bags capable of handling 200 SCFM.

Solid Feeding

Coal/limestone mixtures can be pneumatically fed into the combustor 44.5 cm (17.5") above the distributor plate. The feed rate is regulated using a screw feeder. The whole assembly is mounted on a weighing scale to provide an accurate record of the feed rate.

Gas Sampling and Analysis System

On-line gas analyzers, (CO, NO_x, SO_2, CO_2, O_2) are available to monitor the composition of flue gases. The gas sampling position is located after the gas cooling system prior to entering the secondary cyclone. The sampled gas is conveyed to

the analyzers through a heated line, a gas moisture conditioning system and a diaphragm pump.

Temperature Measurements

Combustor, standpipe and exit gas temperatures are measured with Inconel sheathed K-type thermocouples, located at approximately 30.5 cm (2') intervals along the height of the combustor and the standpipe.

2.3. Project Execution

The project will be divided into two phases. The first phase deals with modification and testing of RPC's existing combustor, while the second phase is to carry out a test program in order to fulfil project objectives as given in Section 2.2.1.

2.3.1 Phase I Tasks

Phase I will be subdivided into five sub-tasks as follows:

2.3.1.1 Feedstock Acquisition, Preparation and Characterization (Task 1.1)

One and a half tons of Minto coal, and 0.5 tons of limestone will be required. Drying, crushing and screening of raw materials will be carried out at RPC.

Characteristics of feedstocks determined will include: particle size distribution, ultimate, proximate and ash analyses.

2.3.1.2 Standpipe Modification and Testing (Task 1.2)

RPC's existing CFBC will be modified in order to reduce the solids capacity of the standpipe. This will be achieved by inserting a 3", 316SS pipe inside the 9" I.D. stand pipe as shown

in Figure 2. This will reduce the solids inventory in the combustor by approximately 75%, with a corresponding reduction in the time required to reach steady state operation.

The effect of standpipe modification on CaS formation, combustion efficiency, bed temperature stability and bed inventory will be evaluated under a range of operating conditions. Three tests each of up to 12h duration will be carried out according to the conditions given in Table 1 (Test 1 to 3). During each test continuous monitoring of bed temperature and gas composition will be carried out. Samples of bed material and baghouse ash will be collected.

The tests will be carried out sequentially, which would allow sufficient time to assess whether loss of bed material is a practical problem. Also the time requirements to reach steady state operation can be assessed from measuring the silica content in the bed.

These three tests will give an indication of the effect of excess air level and air-staging on CaS and NO_x formation and the combustion efficiency.

2.3.1.3 Secondary Air Ports Modification and Testing (Task 1.3)

All the combustion tests performed to date have been carried out with the secondary air ports located 81" above the distributor plate, which represents 40% of the active height of the combustor. This resulted in a reducing zone at the bottom of the combustor which could be a factor in CaS formation.

The secondary air ports will be lowered to an existing set of ports which are located 60" above the distributor plate. Three tests will be carried out in order to assess the effect of this modification on NO_x and CaS formation. The tests to be carried out are in fact a repeat of the earlier tests as shown in Table 1.

Figure 2: Modification To The CFBC

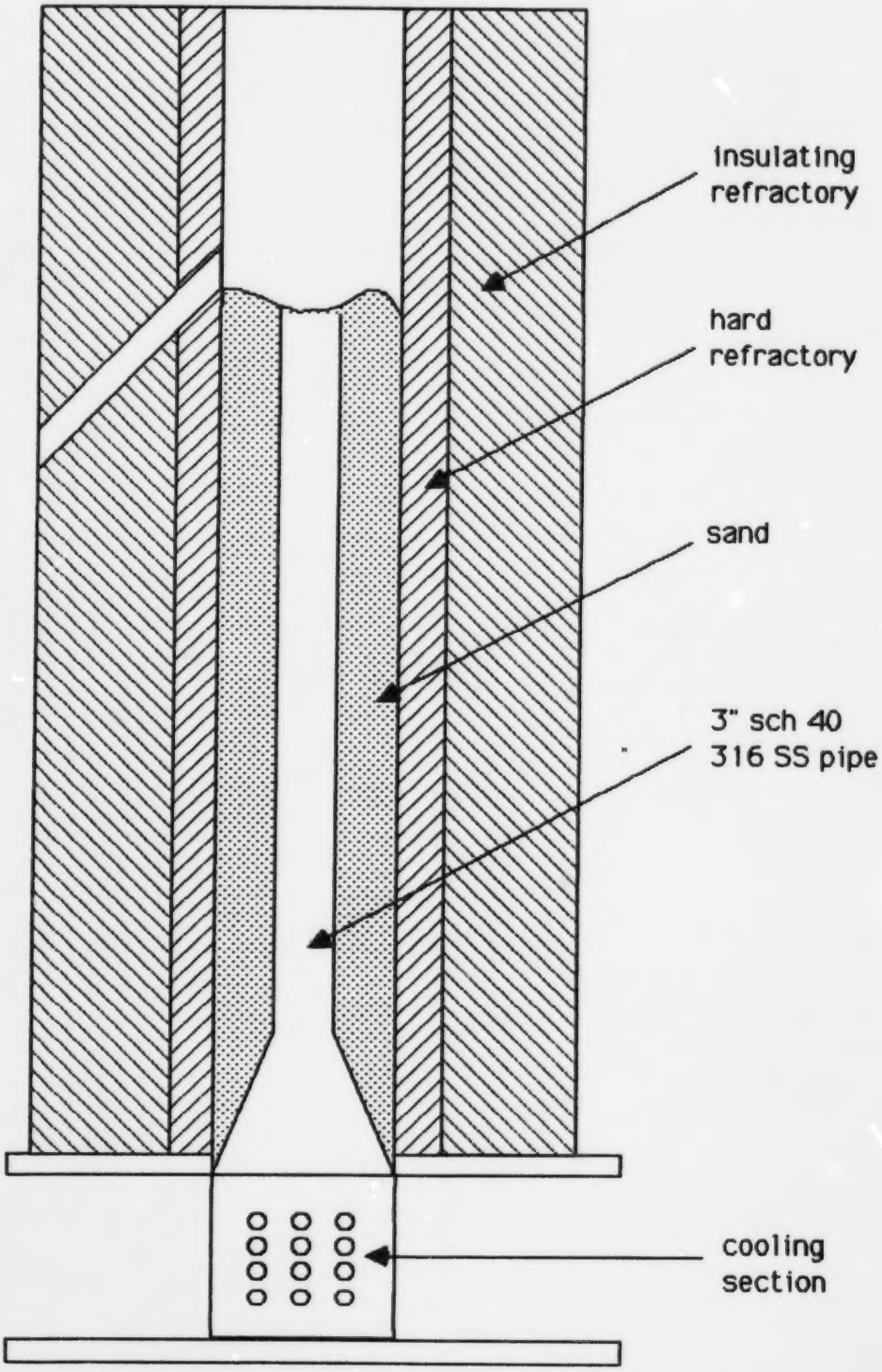

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TABLE 1

Test Matrix for the Assessment
of Combustor Modification

Test No.	Modification	Excess Air %	Air Staging (sec/primary
1	reduce standpipe diameter to 3"	10	40/60
2	"	20	"
3	"	"	25/75
4	lower secondary air ports	"	"
5	"	"	40/60
6	"	10	"

Note: Values for all other operating parameters such as Ca/S ratio, bed temperature, gas velocity will be selected in consultation with the Scientific Authority.

2.3.1.4 Chemical and Data Analysis (Task 1.4)

The product ash streams, i.e. tested material and baghouse ash will be analyzed for total sulphur, CaS and carbon content. This data together with on-line analysis of the flue gases, will be used to calculate combustion efficiency, SO_2 and NO_x emission levels and to assess the degree of CaS formation.

Ash samples will be stored for six months to allow further analysis if necessary.

2.3.1.5 Phase I Report (Task 1.5)

This will be a short report, listing test conditions and results. It will also include an assessment of the effectiveness of bed modifications regarding SO_2, NO_x and CaS formation.

At this stage a meeting will be held with the Scientific Authority and others invited by him to assess Phase I results.

2.3.2 Phase II

This phase will commence upon agreement that representative and meaningful data has emerged from Phase I. It consists of several tasks, the main ones include, testing of Minto and Prince coals with a variety of limestones, assessment of some operating parameters and some test work using oil shale as a sorbent.

2.3.2.1 Feedstocks Acquisition, Preparation and Characterization (Task 2.1)

As mentioned above a variety of coal/sorbent combinations will be tested. There is also the need to correlate the results of this study to that of the demonstration unit at Chatham. For this reason N.B. Power and N.S. Power will be approached for assistance in acquiring the appropriate feedstocks. The estimated requirement of coal are 12 and 3 tons for Minto and Prince coal,

respectively. The total requirement of Havelock limestone (2 sizes) is 2.5 tons, together with 1.0 ton of Elmtree limestone. Oil shale requirment is approximately 10 tons. There are also two other limestones to be tested; the type and quantity will be identified in consultation with the Scientific Authority.

Feedstock preparation may include drying, crushing and screening as appropriate. Most of these tasks will be carried out at RPC. Feedstock characterization will include particle size distribution, ultimate, proximate and ash analyses.

2.3.2.2 Testing of Minto Coal (Task 2.2)

The objective of this task is to test the degree of sulphur capture which can be achieved with Minto coal at different Ca/S ratios and limestones types and particle sizes. Nine tests each of approximately twenty four hour duration will be carried out according to the tentative test matrix shown in Table 2. The final test conditions will be selected in consultation with the Scientific Authority in order to duplicate the conditions of NB Power tests.

For Elmtree limestone the particle size to be used will be that which gave the best results for the Havelock limestone. All three Ca/S ratios will be tested at 100% load and either at 30 or 70% of load. For RPC's combustor operating at constant excess air 100% load means the maximum operating velocity, which is in the range of 8 to 9 m/sec. At 30% load the operating velocity will be as low as 2.4 m/s. This low velocity could lead to operational problems due to slugging, in which case 70% load will be used instead.

The molar Ca/S ratio for test 9 will be selected based on whether sufficient sulphur capture was achieved in test 8.

TABLE 2

Tentative Test Matrix for Minto Coal

Test No.	Type	Limestone Mean particle size, micron	Molar Ca/S ratio	Load %
1	Havelock	500	1.3	100
2	"	"	1.6	"
3	"	"	2.0	"
4	"	150	Optimum	"
5	Elmtree	Optimum	1.3	"
6	"	"	1.6	"
7	"	"	2.0	"
8	"	"	1.6	30 or 70
9	"	"	2.0 or 1.3	"

Other operating parameters such as bed temperature, excess air, air distribution, etc. will be selected to duplicate conditions run by NB Power and after consultation with the Scientific Authority.

2.3.2.3 Chemical, Data Analysis and Reporting (Task 2.3)

Chemical analysis will be performed on bed material*, and bed and baghouse discharges for unburned carbon, calcium species, total sulphur and sulphides. Trace element analysis will be carried out on approximately 20% of the samples collected, to be selected after consultation with the Scientific Authority. CANMET will be provided with a full set of samples from all tests. These analyses in addition to flue gas analyses, will be sufficient to calculate parameters such as calcium utilization (% SO_2 reduction) and combustion efficiency.

The other objective of data analysis will be to compare the test results with those from the equivalent tests performed on the Chatham unit. An interim report will be compiled and a presentation will be given in Fredericton to interested parties who will be invited by the Scientific Authority. At this stage the work will be evaluated by the Scientific Authority in order to provide direction for the remainder of the project.

2.3.2.4 Testing of Prince Coal (Task 2.4)

In this task six combustion tests each of approximately 24 hour durations, will be carried out using Prince coal. The limestones to be tested and their particle sizes will be selected upon consultation with the Scientific Authority. A tentative test matrix is given in Table 3. These tests will be sufficient to rank the limestones tested in terms of their capacity for sulphur capture.

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* For low ash fuels, bed discharge is negligable i.e., all the ash in the feed streams reports to the baghouse. In extreme cases there even could be a loss in bed inventory to the baghouse. Under these conditions bed discharge from RPC combustor is not representative of bed material. This phenomena is rig specific and is probably due to collecting bed discharge as an overflow through a side port located near the top of the standpipe (i.e. near the bottom of the primary cyclone). Some of the fine particles of dust could find there way into the side port and report as bed discharge. For high ash fuels however, particularly for oil shale, bed discharge is significant and is probably representative of bed material.

TABLE 3

Test Matrix For Prince Coal

Test No.	type*	Limestone new particle size, micron	Ca/S Ratio
1	A	500	2.0
2	A	150	2.0
3	A	optimum	1.5
4	B	"	2.0
5	B	"	1.5
6	C	"	Optimum

* Limestone types to be selected upon consultation with the Scientific Authority.

2.3.2.5 Chemical, Data Analysis and Reporting (Task 2.5)

This task will be similar in scope and objectives to Task 2.3 given above. The results will be compiled into an interim report to be submitted to the Scientific Authority. At this point another presentation will be given in Fredericton to a group invited by the Scientific Authority.

2.3.2.6 Testing of Operating Parameters (Task 2.6)

There are a number of operating parameters which could have a significant effect on SO_2 and NO_x emission as well as CaS formation. In this study however, the effect of air distribution will be examined. Five tests each of approximately 24 hours duration will be carried out according to the test matrix given in Table 4.

The tests will be carried out using Minto coal and Havelock limestone. The rest of the operating conditions will be the optimal value as determined in Task 2.2.

2.3.2.7 Chemical and Data Analysis (Task 2.7)

Chemical analysis will be similar to those given in Task 2.3. Date analysis will consist of identification of the effect of air distribution on combustion efficiency, SO_2 capture, CaS formation and NO_x emission.

2.3.2.8 Oil Shale Tests (Task 2.8)

The effect of oil shale on the combustor performance in terms of combustion efficiency and pollutant emission will be investigated. The two variables to be tested are oil shale size distribution and Ca/S ratio. Six tests are to be performed according to the test matrix given in Table 5. Coal/oil shale ratio will be fixed at 1 to 2.2, which to the best of our knowledge, is the proportion to be used at Chatham. Test one will

provide a base case for sulphur capture. In the subsequent tests an appropriate amount of limestone will be added to bring the Ca/S ratio to the desired level. The remainder of the operating conditions will be fixed at selected values to be determined in consultation with the Scientific Authority.

2.3.2.9 Chemical and Data Analysis (Task 2.9)

Chemical analysis will be similar in scope to those given in Task 2.3. Additional analysis will be carried out in order to qualify the degree of magnesium utilization, i.e. the amount of $MgSO_4$ in the waste streams. Data analysis will show whether oil shale particle size has a significant effect on sulphur capture.

2.3.2.10 Final Report (Task 2.10)

The testing procedure used to obtain the experimental results will be given together with the test conditions and results. The correlation between RPC's pilot plant performance and the Chatham unit will be present. Recommendations regarding the suitability of tested limestones and oil shales will be given. The report will also include recommendations for future work.

TABLE 4

Test Matrix For Assessment Of
The Effect Of Air Distribution

Test No.	Air Distribution pri/sec ratio
1	50/50
2	60/40
3	40/60
4	min. primary feasible
5	reverse of Test 4

Note: The remainder of operating conditions will be the optimum determined in Task 2.2

TABLE 5

Test Matrix For
Oil Shale Testing

TEST NO.	Ca/S Ratio	Oil Shale* Particle Size
1	-	S
2	1.3	"
3	1.6	"
4	2.0	"
5	Optimum	N
6	"	B

* Particle size distribution

S = Small (18 mesh to zero)
N = Narrow (6 mesh to 18 mesh)
B = Broad (6 mesh to zero)

3.0 MANAGEMENT PROPOSAL

3.1 Management and Personnel

The project will be undertaken by the Fuel Technology Section of the Research and Productivity Council. The overall management of the program will be by P. Salib, Head, Fuel Technology, under M. Chalkley, Head of Mineral Development and Processing. R. Robson, Chemical Engineer, will provide technical assistance. Other RPC professionals and technicians will be used as necessary. Résumés of key personnel are attached in Appendix C.

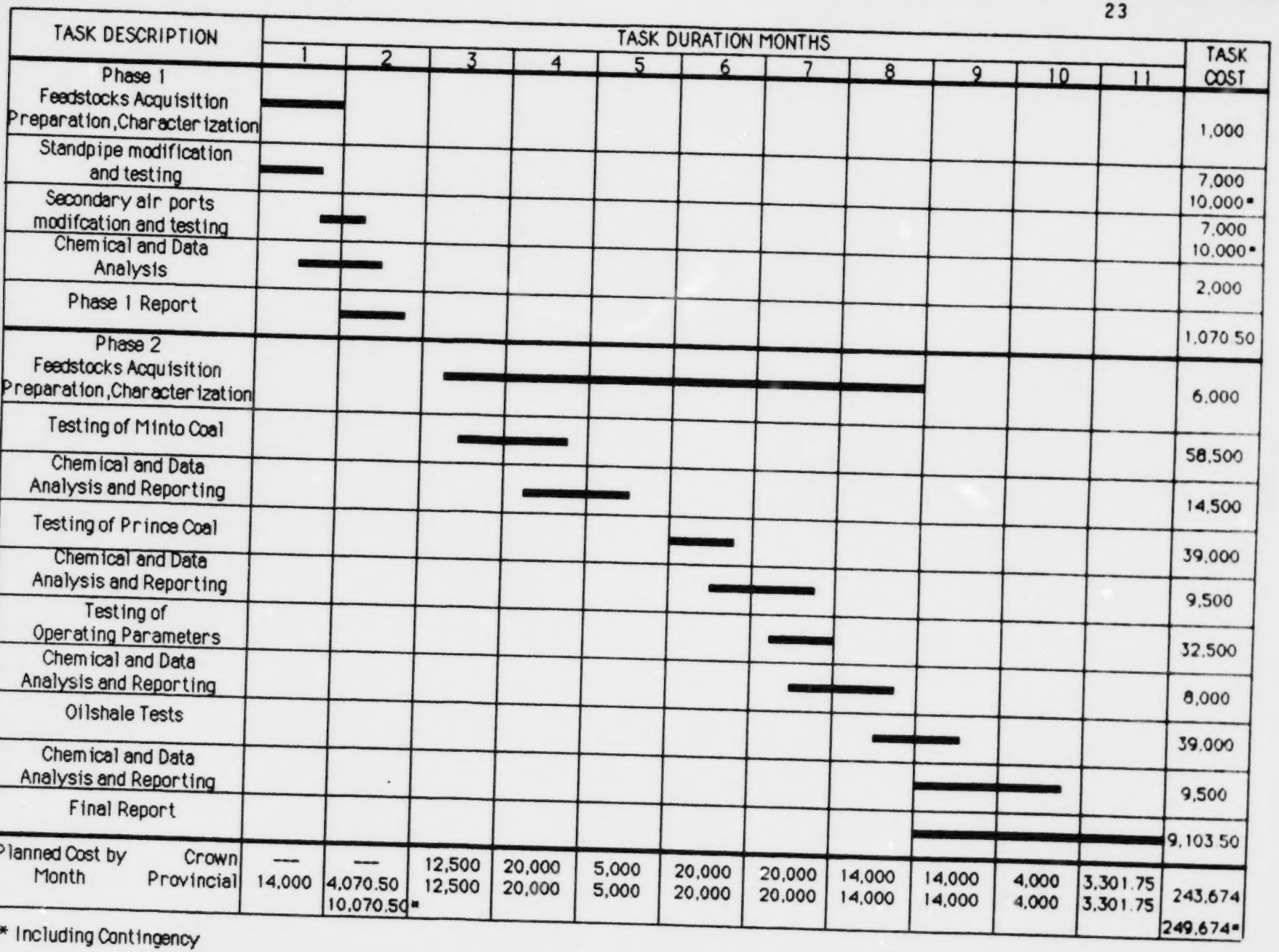
3.2 Project Schedule

The project can be completed within eleven (11) months of commencement.

A tentative time schedule for each task as well as estimated costs are given in Figure 3.

3.3 RPC's Credentials

RPC is at the forefront of circulating fluidized bed technology in Canada. Among the recently completed and on-going projects involving the circulating fluidized bed are:



* Including Contingency

Figure 3: Time and Cost Schedule

1986 Circulating fluidized bed combustion of 10 different coals/limestones combination for Dearborn Chemicals Ltd on behalf of CEA. The main objectives of the work were to produce CFB ash under realistic operating conditions for subsequent characterization for environmental impact.

1985 Pyrolysis of Eastern Canadian Coals and Oil Shales in a Circulating Fluidized Bed.

Phase 1: Construction of the unit. The project was completed in August 1985. DSS has recently issued a contract for continuation of the project.

1985 Modification of Integrated Oil Shale Retort System. This is an on-going project to develop the integrated system. It is expected to be completed in March 1987. Oil shale is retorted to recover oil and gas, and the spent shale is burnt with the high sulfur coal in the circulating fluidized bed combustor to produce heat for retorting. Sulphur trapped by the spent shale reduces emission to the atmosphere.

Other fluidized bed combustion projects completed at RPC were:

1981-82 Fluidized bed combustion tests on Minto coal and Albert County oil shale which demonstrated the capability of the oil shale to capture sulfur. These findings led the Federal Government and New Brunswick Electric Power Commission to undertake a $3million project for the demonstration of CFBC technology at Chatham, N.B.

1982-83 Design of processing concepts on co-utilization of high sulfur coal and oil shales in the integrated system of combustion and retorting. Retorting of low grade oil shale was carried out in 9" x 9"

retort for the production of liquid and gaseous hydrocarbon.

1983-84 Development of integrated system. Retorting of better grade N.B. oil shale was carried out in 9" x 9" retort. The spent shale was burnt with high sulfur coal in a fluidized bed. Spent shale was shown to have good sulfur capture capability.

RPC has previously been funded under the cost shared program for projects to the value of $424,461. The results of this work has played a major part in the development of oil shale technology. The projects are as follows:

Sponsorship CANMET, Ottawa, and Provincial Holdings Ltd. New Brunswick under CANMET Energy Conversion Cost Shared Program

1983/84	DSS 15SQ	$253,342	Demonstration of an Integrated Oil Shale Retort System
1985/86	DSS 06SQ	72,928	Modification of the Integrated Oil Shale Retort System
1985/86	DSS 0SQ-0035	98,141	Pyrolysis of Eastern Canadian Coals and Oil Shales in a Circulating Fluidized Bed reactor. Phase 2 Unit Operation.

4.0 PRICE PROPOSAL

The total expenditure anticipated for this project is $249,674.00 which includes $6,000.00 contingency. The costs are to be divided between the Federal and Provincial Governments as shown in Table 6. The project work plan by task and month including approximate cost is outlined in Figure 3. Labour costs are based on 86/87 DSS/RPC negotiated rates escalated by 3%. A summary of costs are given in Table 6 and a cost break down for equipment, material and supplies are given in Table 7.

TABLE 6
Summary of Costs

1. Direct Labor Charges (RPC labor rates are those negotiated with DSS for FY 86/87 escalated by 3%).

Category		Rate $/hr.	Total Hours Phase I	Total Hours Phase II	Cost ($) Phase I	Cost ($) Phase II	Total Cost ($)
Professional	P5	89.00	5	100	445.00	8,900.00	9,345.00
	P4	83.40	20	340	1,668.00	28,356.00	30,024.00
	P3	71.20	40	390	2,848.00	27,768.00	30,616.00
	P2	54.50	40	480	2,180.00	26,160.00	28,340.00
	P1	45.60	40	470	1,824.00	21,432.00	23,256.00
Technician	T3	62.30	30	100	1,869.00	6,230.00	8,099.00
	T2	44.50	60	750	2,670.00	33,375.00	36,045.00
	T1	28.90	60	750	1,734.00	21,675.00	23,409.00
Lab Assistant	LT2	34.50	25	760	862.50	26,220.00	27,082.50
	LT1	24.50	60	775	1,470.00	18,987.50	20,457.50
			TOTAL DIRECT LABOR		17,570.50	219,103.50	236,674.00

2.0 Direct Charges at laid down costs without markup			
Material and supplies (see Table 4)	500.00	5,000.00	5,500.00
Drafting	--	500.00	500.00
Travel	--	1,000.00	1,000.00
TOTAL DIRECT CHARGES	500.00	6,500.00	7,000.00
3.0 Contingency	6,000.00	--	6,000.00
TOTAL COSTS	24,070.50*	225,603.50	249,674.00*

4.0 Cost Distribution	
Crown Share (50% of Phase II costs)	$112,801.75
Provincial Government Share (Phase I costs + 50% of Phase II)	136,872.25*

* Including $6,000 contingency.

TABLE 7

Material and Supplies

Item	Estimated Cost
Standard gases for the gas analysis	$2,000
Thermocouples	1,000
Safety supplies	500
Stainless steel pipe, sheets and fittings	500
Miscellaneous supplies	1,500
	$5,500

5.0 CERTIFICATION

Price

"We hereby certify that the price quote is based on costs computed in accordance with Costing Memorandum DSS 1031, copy of which we acknowledge to have received, and does not contain any element of profit."

Foreign Content

"The value of the foreign content in our price is $ nil Canadian funds. This value does not include the cost of processing materials of foreign origin in Canada by us or by our subcontractors or any element of our overhead or profit, duties, sales tax and transportation in Canada".

Cost Guidelines

The quoted contract prices is fixed (e.g., ceiling price contract). The hourly rates quoted are firm for the duration of the contract.

AUTHORIZATION

DEPARTMENT: APPLIED SCIENCES

PROJECT TITLE: INVESTIGATION OF SULPHUR CAPTURE DURING CIRCULATING FLUIDIZED BED COMBUSTION OF EASTERN CANADIAN COALS

PROJECT NUMBER: PROPOSAL NUMBER AS/87/30P (revised)

REPORT

Prepared By: P. Salib
R. Robson

Reviewed By: M. Chalkley

Approved By:

Date: April 10, 1987

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APPENDIX A

SAMPLE CALCULATIONS

Steady State Bed Composition Using Devco Coal and Elmtree Limestone

ASSUME 1) 100% coal combustion
2) complete sulphur capture
3) no $CaCO_3$ in the discharge

Calculate Bed Composition:

BASIS 100 kg coal feed

Ca/S = 2.5 Sulphur in the coal = 3.41

$CaCO_3$ required = $\frac{3.41}{32} \times 2.5 \times 100 = 26.64$ kg of $CaCO_3$

ELMTREE is 93.23 % $CaCO_3$

. . needed 28.57 kg of Elmtree

$CaCO_3$ unutilized = $\frac{56.08}{100.08} \times \frac{1.5}{2.5} \times 26.64 = 8.96$ kg $CaCO_3$

$CaSO_4 = \frac{136.14}{100.08} \times 10.66 = 14.50$ kg

ASH = 11.6 kg

. . Comp = CaO 24.2%
$CaSO_4$ 39.2%
Ash 31.4%
Other 5.2%
100 %

Actual Ca/S Ratio in the Combustor

BASIS: Devco + Elmtree

Ca/S ratio in the feed = 2.5

From Table 1 the CaO in the bed = 24.2%

ASSUMING 4) Recycle rate = 2000 kg/hr
5) Coal feed rate = 10 kg/hr

. . Recycle ratio = $\frac{2000}{10}$ = 200

Sulphur in the coal = 0.0341 $\frac{kg}{kg\ coal}$

Calcium oxide in the recycle

= 200 x 0.242 = 48.4 $\frac{kg}{kg\ coal\ feed}$

Aparent Ca/S molar ratio

$$= \frac{\text{mass of CaO in the recycle}}{\text{molecular weight of CaO}} \times \frac{\text{molecular weight of S}}{\text{mass of S in the coal feed}}$$

$$= \frac{48.4}{56.08} \times \frac{32.06}{0.0341} = 811$$

APPENDIX B

RELEVANT WORK CARRIED OUT AT RPC

ABSTRACT

The combustion of bituminous coals of varying sulphur content and a lignite coal was carried out in a pilot scale circulating fluidized bed combustor (CFBC). High calcium limestones were used for sulphur capture ranging from 60 to 97% at Ca/S molar ratios of 1.5 to 2.5:1. The combustion efficiencies achieved varied from 88 to 99%. NO_x emission was very low in the range of 88 to 219 ppm. CaS was present in appreciable concentration in many of the waste samples tested. Its formation is probably related to air/fuel ratio, air distribution and secondary air injection location.

INTRODUCTION

The combustion of a range of coal/limestone combinations from Canada and the United States was carried out in a pilot scale circulating fluidized bed combustor (CFBC) at the Research and Productivity Council (RPC). The main objective of the study was to generate representative solid residues for full characterization of their physical, chemical and engineering properties in order to identify the potential impact of their disposal (1). In this article however, the emphasis will be on some of the combustion aspects of the work.

The coals burnt were three bituminous high sulphur coals and a lignite. Sulphur content varied from 0.6 to 6%. All three limestones used were high in calcium, however there was a wide variation in their particle sizes and possibly in their reactivities.

The study has highlighted some of the significant parameters affecting the degree of sulphur capture and identified a potential problem of CaS formation which could have serious implication for waste disposal and potential use. Preliminary testing shows that CaS formation can be minimized by appropriate design modification and careful selection of the operating conditions.

TEST FACILITY

The Research and Productivity Council's circulating fluidized bed pilot-scale test facility is shown schematically in Figure 1. The main components of the system are the combustor, a primary cyclone and a standpipe for solid recirculation, a cooling section, a secondary cyclone and a baghouse.

The refractory lined combustor is 127 mm in diameter and has an overall height of 5.8 m. It is equiped with two sets of secondary air ports located at 1.54 m and 2.05 m above the distributor plate which is located near the bottom of the combustor. The latter ports were the ones used in this particular study.

Solid recirculation is carried out using a primary cyclone, a standpipe and a J-valve. The primary cyclone and the standpipe are refractory lined and are 0.3 and 0.23 m in diameter, respectively. The latter is 2.74 m in height. A cooling section is attached to the bottom of the standpipe. It has 21 concentric tube cooling lances each with an outside diameter of 15.8 mm. The lances are arranged in seven banks which can be controlled independently. The cooling section is used to control bed temperature as well as giving an indication of the solid recirculation rate. This is deduced from a simple energy balance on the heat loss by the recycled solids versus the heat gained by the cooling water. The circulation rate is controlled by adjusting the J-valve's aereation and lift air flows.

The flue gases leaving the primary cyclone are cooled in a water-jacketed 6" pipe, 5.2 m in height. This is followed by a stainless steel 0.25 m diameter secondary cyclone. The gases are finally fed into a self cleaning dust collector which consists of 16 high temperature bags. In this particular study all the solids collected by the secondary cyclone were recycled back into the combustor.

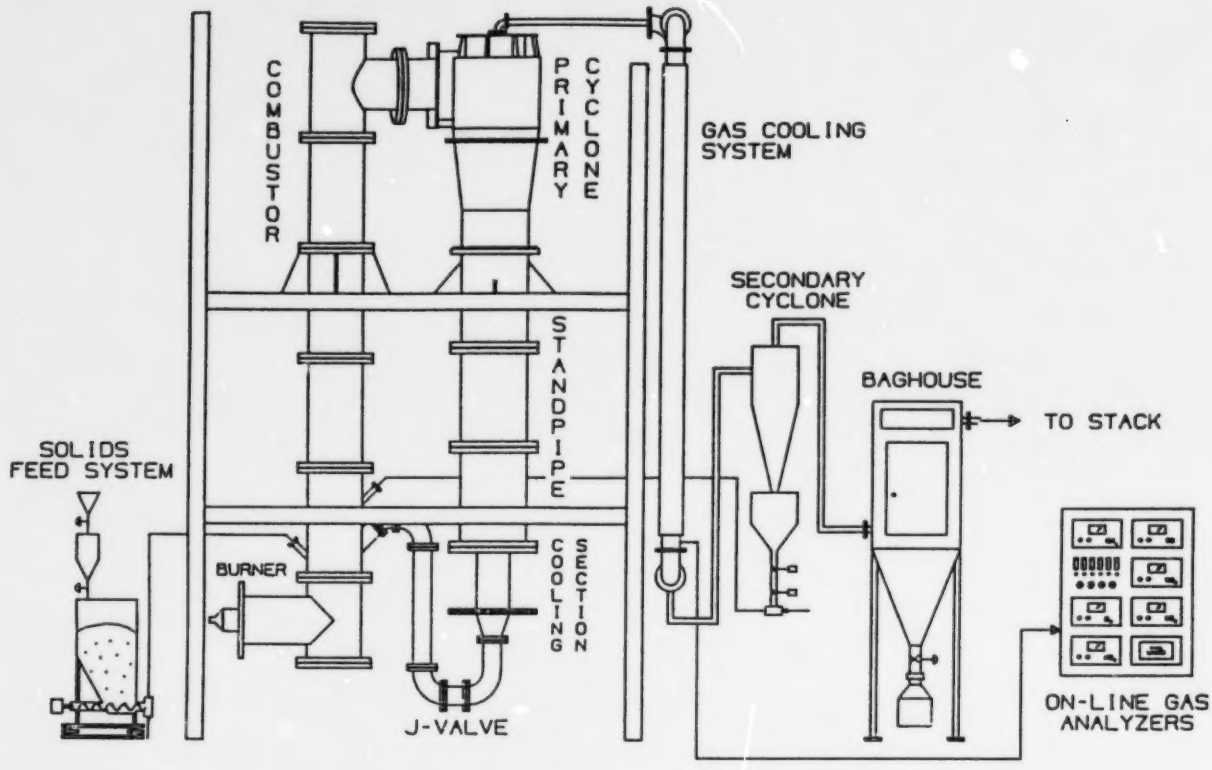

Fig 1: RPC circulating fluidized bed combustor

Other auxiliary equipment includes a 0.5 million Btu/hr propane burner for start up. A solid feeding system, which is a simple screw feeder mounted on a weighing scale to provide an accurate account of the feed rate. The feed material is then pneumatically fed into the combustor. The system is also equipped with on-line gas analyzers for CO, CO_2, O_2, NO_x and SO_2. A mirco processor is also available for data-logging of gas analysers outputs and for temperature measurements. The system is also equipped with 32 channel data logger.

EXPERIMENTAL PROCEDURE

For all the experimental runs except for test no. 11, as given in Table 1, the combustor was started on a fresh bed of approximately 250 kg of silica sand of a top size of 0.6 mm. This was necessary in order to generate the different solid wastes which was the main objective of the study (1). The combustor was then preheated to a temperature of approximately 600°C using a 0.5 M Btu/hr propane burner. At this temperature the coal feed was initiated and the burner was switched off. Once the bed temperature reached 850°C, the cooling section at the bottom of the standpipe was activated and the operating conditions were adjusted for the particular test. From a cold start this procedure requires approximately 24 hours to complete. The combustor was then operated for a minimum of 2 hours to ensure stable bed temperature and flue gas composition. Only then was a test run commenced, which was then continued for a minimum of 12 hours.

For all tests the experimental measurements consisted of the following:

(i) temperature measurements along the length of the combustor and standpipe as well as at several other important locations such as the primary cyclone exit, baghouse inlet, etc.

(ii) On-line gas analysis for CO, CO_2, O_2, SO_2 and NO_x

(iii) All input air flow rates and coal-limestone mixture feed and solid discharge rates from the bed and the baghouse.

Upon completion of a test run, samples from the bed discharge and baghouse ash were fully characterized in terms of chemical and physical properties. Some of these properties are shown below and more detailed analyses can be found elsewhere (1).

For test no. 11, the combustor was started on silica sand and limestone at a ratio of 70:30. This amount of limestone was estimated to provide 75% of the bed equalibrium CaO concentration, which was

estimated to be approximately 24%. This procedure was adopted in order to reduce the time requirement to reach steady state bed composition, i.e., a bed free from the sand used as starting bed material. Periodic sampling of the bed material during this test confirmed that the CaO concentration was rising and approaching an equilibrium (2).

RESULTS AND DISCUSSION

The operating conditions and the test results are shown in Table 1. The properties of the three bituminous coals and the lignite coal (Sask.) tested are shown in Table 2, while that for the limestones are given in Table 3.

Most of the parameters shown in Table 1 are self-explanatory, however, there are some which deserve comment. Air was distributed as a primary stream fed at the base of the combustor, a secondary stream fed 2.05m above the distributor plate, which corresponds to 40% of the bed height and "others". The latter refers to pneumatic transport air for the coal/limestone feed and recycled solids and air for the operation of the J-valve. All these streams entered the combustor just above the distributor plate. Therefore, for all practical purposes, these steams have the same effect as the primary air stream. Sulphur reduction was calculated on the assumption that coal ash does not retain any sulphur. The efficiency was calculated on the basis of carbon in the feed and that in the flue gases (gas basis) or carbon loss in the ash (solid basis). The results given in Tables 1 and 4 will be discussed in terms of combustion efficiency, NO_x and SO_2 emission and CaS formation.

Combustion Efficiency

The combustion efficiencies achieved were in the range of 88 to over 99%. The lignite (Sask), not surprisingly, gave the highest combustion efficiency.

These results are good for a pilot-scale unit where the residence time in the combustor is less than 1 second which is much shorter than what would be expected in a commercial size unit. Also, the excess air levels for most of the experimental runs were quite low.

Nitrogen Oxides Emission

Nitrogen oxides emissions were in the range of 88 to 219 ppm which is what would be expected from a circulating fluidized bed operating at low excess air levels and with a relatively high degree of air staging. For test no. 12 however, there was no secondary air stream i.e., all the air requirements were fed as primary and "others" which, as explained above, can be considered as a primary air stream. The surprising result is that NO_x emission remained low. These emission levels would meet the most stringent environmental standards.

TABLE 1
Operating Conditions and Test Results

Parameter	Test No 1	2	3	4	5	6	7	8	9	10	11	12
Coal Type	Devco	Devco	Devco	Sask	Sask	Sask	US	US	US	Minto	Devco	Devco
Batch	2	2	1	1	1	2	2	2	1	1	3	4
Sorbent Type	HL	Ex	Elm	HL	Ex	Elm	HL	Ex	Elm	HL	Elm	Elm
Ca/S (molar ratio)	1.5	2.0	2.5	2.0	2.5	1.5	2.5	1.5	2.0	2.4	2.5	2.0
Bed Temp °C	846	848	840	837	851	846	850	847	840	840	832	857
Velocity ft/s	26.6	25.4	27.0	23.8	25.1	25.3	27.3	24.8	25.2	28.2	22.3	21.3
Feed rates (lb/hr)												
Coal	28.31	27.06	25.53	30.78	33.41	36.7	26.08	22.62	23.07	28.65	22.89	20.26
Sorbent	5.44	6.64	7.28	1.29	1.72	1.17	3.56	1.91	2.36	13.61	6.82	5.19
Air	274.04	260.6	278.9	247.2	256.9	260.6	279.9	255.0	260.6	291.8	233.3	206.5
Air Distribution %												
Primary	44.8	38.5	40.2	35.2	39.1	38.5	40.1	39.3	38.5	42.1	30.5	74.1
Secondary	27.2	31.3	31.6	33.0	31.8	31.3	29.2	32.0	31.3	34.2	33.6	0.0
Others	28.0	30.2	28.2	31.8	29.1	30.2	30.7	28.7	30.2	23.7	35.9	25.9
Theoretical Excess Air %	7.56	6.88	19.4	18.3	13.3	7.00	4.98	10.3	8.44	1.69	25.3	13.2
Discharge Rate (lb/hr)												
Bed	0.18	0.38	0.42	0.36	0.21	0.29	0.32	0.39	0.46	1.47	1.34	0.48
Baghouse	6.58	4.67	3.57	2.09	2.96	4.6	7.72	3.06	4.11	8.04	9.33	8.82
Gas Composition (dry basis)												
O_2 %	2.94	3.17	3.18	3.5	3.6	2.97	3.23	3.18	3.34	2.78	3.70	3.84
CO_2 %	15.74	15.33	16.05	16.61	16.0	16.79	15.05	15.51	15.11	16.76	15.27	14.54
SO_2 ppm	1601	1420	549	211	315	246	428	517	360	976	87	882
NO_x ppm	116	134	88	133	157	99	104	93	97	110	219	170
Sulfur Red. %	57.5	62.2	80.5	70.6	57.9	66.6	69.8	61.5	71.0	81.6	97.3	74.4
Comb Efficiency %												
Gas Basis	92.6	90.7	101	98.4	93.9	92.1	89.2	93.9	91.6	88.2	103	88.3
Solid Basis	88.7	93.7	96.7	99.6	99.1	99.3	91.1	95.1	95.1	95.3	-	89.2

HL = Havelock Ex = Exshaw Elm = Elmtree

TABLE 2
Chemical and Size Analysis of Coals Tested

Parameter	COAL TYPE								
	Devco				Sask		US		Minto
Batch No.	1	2	3	4	1	2	1	2	1
Chemical Analysis wt%									
Moisture	1.2	0.29	0.1	-	5.34	14.1	0.77	0.11	0.57
Ash	11.60	17.1	20.76	17.13	11.1	18.1	7.87	7.7	14.2
Volatile Matter	35.9	35.4	-	-	48.2	35.9	36.5	39.8	35.6
C	68.33	65.31	60.43	65.4	54.87	49.40	77.11	76.53	71.57
H	4.67	4.92	4.39	4.94	4.32	4.44	5.23	5.05	4.84
N	1.19	1.11	1.14	1.12	0.66	0.51	1.51	1.24	0.94
S	3.41	4.01	3.57	3.81	0.64	0.57	1.53	1.65	6.0
O	9.60	7.27	9.73	7.31	23.1	12.94	6.0	7.72	1.95
Calorific Value Btu/lb	-	11550	-	-	-	7596	13686	13575	12855
Sieve Analysis									
+10 mesh	16.6	19.8	16.4	-	37.7	28.5	25.4	21.1	37.7
-10 +20	25.3	28	29	-	27.5	26.9	27.0	28.5	26.5
-20 +30	15.7	14.8	15.9	-	14.7	12.2	12.5	13.9	11.7
-30 +40	6.4	6.0	6.4	-	3.3	4.8	5.3	5.6	4.3
-40 +100	22.9	16.7	17.8	-	11.0	13.6	15.0	15.2	11.6
-100 +200	7.3	6.1	6.2	-	3.5	6.0	8.5	5.7	3.8
-200 +300	3.9	2.8	2.7	-	1.2	3.5	5.0	2.5	1.9
-300 +350	0.4	5.9	5.6	-	0.1	4.4	1.3	7.5	2.4
-350	1.4	-	-	-	1.0	-	-	-	-

Sulphur Dioxide Emission

The results given in Table 1 and Figure 2, show that sulphur reduction increases as the Ca/S ratio is increased. However, with the exception of run no. 11, the degree of sulphur reduction achieved was relatively low for a circulating fluidized bed combustor. Some explaination of the causes are discussed below.

Figure 2 shows that among the three limestones tested Elmtree was the most effective followed by Havelock then Exshaw. This result however should be interpreted with caution because of the differences in the particle size of the limestone used, particularly Exshaw which, as shown in Table 2, consisted of a narrow size range with no fine fraction.

The operating conditions for test No. 11 is similar to that of test no. 3, with the exception that the starting bed material for test No. 1 was a 30:70 mixture of limestone and sand. Comparing the sulphur reduction for these two tests show a significant improvement from 80.5% to 97.3%. The importance of changes in bed composition becomes apparent considering that the amount of recycled solids employed in this study were two orders of magnitude higher than the feed rate. Therefore for test No. 11 the actual molar Ca/S ratio was approximately 800:1 compared to 2.5:1 in the feed stream. For test no. 3 however, the recycled solids was mostly inert sand.

These findings highlight the significance of the recirculation rate as an important operating parameter for sulphur control. The recirculation rate may also have a significant impact on the combustion efficiency, bed pressure drop and overall boiler efficiency.

TABLE 3
Chemical and Size Analysis of Limestones

Paramter	Limestone		
	Havelock	Exshaw	Elmtree
Chemical Analysis(wt%)			
Calcium	36.2	37.2	37.34
$CaCO_3$	90.39	92.89	93.24
Magnesium	0.59	0.76	0.27
$MgCO_3$	2.04	2.63	0.94
Sieve Analysis			
+10 mesh	-	-	-
-10 +20	15.5	70.5	8.0
-20 +50	47.0	29.5	40.0
-50 +100	23.5	-	31.0
-100	13.5	-	21.0

Calcium Sulphide Formation

The results given in Table 4 show that, with few exceptions, CaS concentration in the bed and baghouse materials is high. A reducing atmosphere is conducive to CaS formation. The two reducing areas in this unit are the standpipe and the lower section of the combustor. High concentration of CaS in the bed material as compared to the baghouse suggests the possibility of CaS formation within the standpipe. One possible reaction mechanism for its formation in the standpipe is the reduction of $CaSO_4$ to CaS in the presence of carbon (3).

$$CaSO_4 + 2C = CaS + 2CO_2$$

The second possible explanation, and the one favored, is that CaS forms predominantly in the reducing zone of the combustor which for this tests represent 40% of the bed height. The higher CaS concentration in the bed as compared to the baghouse can be particle size related. Coarser particles are preferentially retained by the primary cyclone and may remain longer in the reducing zone of the combustor hence the high CaS concentrate in the bed material as compared to the baghouse.

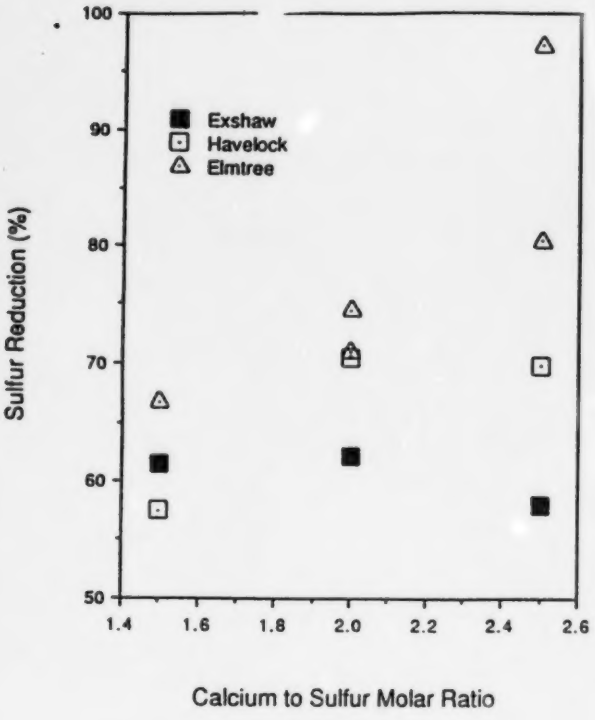

Fig 2: Effect of calcium to sulfur molar ratio on sulfur reduction with various limestones

Test run No. 12 was carried out specifically as a first attempt to verify that CaS forms in the reducing zone of the combustor. The test was conducted with no secondary air stream i.e. there was no reducing zone in the bottom of the combustor. The operating conditions for test no. 12 are shown in Table 1. With the exception of air distribution, the operating conditions are similar to those for tests 3 and 11. However, the CaS levels given in Table 4 are significantly lower for test No. 12 products as compared to that for tests 3 and 11. While these results indicate that the formation of CaS is mostly in the combustor and not the standpipe. It is only a single test run and clearly further testing is required to gain a better understanding of this phenomena.

CONCLUSIONS

Different ranking coals were burnt with varying degrees of combustion efficiency, with the highest being for low ranking coals. The combustion efficiencies achieved were high for a pilot scale unit and would certainly improve on scale up.

The high recycle rate of solids in a circulating bed increases the actual population of active sulphur capture species (i.e. Ca/S ratio) by at least 2 orders of magnitude compared to that in the incoming feed. This results in the high sulphur capture which is typical of circulating fluidized beds.

TABLE 4
Chemical Composition of Bed and Baghouse Samples

Sample*	Chemical Species, wt% CaS	$CaSO_4$	$CaCO_3$	CaO	Carbon
Bed-1	5.3	4.2	2.0	9.3	-
BH-1	0.4	8.8	1.8	14.6	30.8
Bed-2	1.8	0.8	7.1	1.0	0.84
BH-2	1.1	6.4	1.4	1.9	22.2
Bed-3	4.8	5.9	5.7	10.2	2.78
BH-3	1.3	16.7	1.6	15.2	15.0
Bed-4	3.0	2.4	2.5	15.1	0.79
BH-4	0.0	14.9	2.3	20.0	2.72
Bed-5	1.4	0.7	2.5	5.7	0.62
BH-5	0.0	12.1	5.0	11.0	5.4
Bed-6	0.9	1.3	0.2	8.1	-
BH-6	0.3	7.4	6.4	9.8	2.6
Bed-7	1.0	2.5	2.3	6.4	-
BH-7	1.0	8.2	4.8	9.4	22.1
Bed-8	1.4	2.6	8.9	0.7	1.05
BH-8	1.0	4.2	1.8	1.9	24.7
Bed-9	2.7	4.8	2.5	10.6	1.75
BH-9	1.0	8.9	2.3	5.6	19.2
Bed-10	2.5	0.7	2.3	5.5	0.75
BH-10	6.3	8.4	4.3	36.0	10.2
Bed-11	3.8	6.7	2.7	14.6	1.27
BH-11	2.6	15.0	2.3	21.0	-
Bed-12	0.02	-	-	-	0.12
BH-12	0.05	-	-	-	16.9

* Bed-1 - Test 1 bed material
BH-1 - Test 1 baghouse ash

Air distribution and excess air levels affect NO_x emission as well as CaS formation. Under certain operating conditions CaS can be present with high concentration in the waste product streams, in particular in the bed material. Therefore careful consideration should be given to combustor design and the selection of operating conditions to minimize both gas pollutants and the undesirable species in the waste solid streams.

ACKNOWLEDGEMENTS

The authors gratefully acknowledge the financial support of the Canadian Electrical Association and the helpful discussion with Dr. E.J. Anthony, Research Scientist, CANMET.

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2. Robson, R., Salib, P., Ritchie, C. and Barua, S., "Circulating Fluidized Bed Waste Production" RPC Report No. AS/87/10, June 9, 1986.

3. Anthony, E.J., CANMET, personal communication, Jan. 1987.

APPENDIX C

RESUMES

M. Chalkley

P. Salib

R. Robson

RPC - MINERAL DEVELOPMENT AND PROCESSING

MICHAEL EDWARD CHALKLEY, B.Sc., Ph.D., C.Eng.
Head, Department of Mineral
Development and Processing

EDUCATION

1969 - 1972	University of Newcastle upon Tyne studying for a Ph.D.
1974	Awarded Ph.D., Title of Thesis: Gas Injection Phenomena in Water and Mercury.

EXPERIENCE

October 1983 - Present	Senior Hydrometallurgist, BP Research Centre, Sunbury on Thames, Middlesex, England. Responsible for all zinc research carried out by the group. Advisory role for all hydrometallurgical testwork.
April 1980 - September 1983	Senior Research Metallurgist, Sherritt Gordon Mines Limited, Fort Saskatchewan, Alberta, Canada. Responsible for the laboratory testwork for the development of a pressure leach process for the treatment of New Brunswick complex sulphide concentrate. Responsible for the design, construction and commissioning of continuous pressure leaching and liquid-solid separation equipment on a laboratory scale. Assisted in several continuous pilot studies on zinc, cobalt, nickel and gold concentrates.
February 1978 - March 1980	Research Associate, University of Newcastle upon Tyne. Employed on a Science Research Council contract, studying Tuyere Mouth Erosion by physical modelling.
November 1974 - November 1977	Matthey Rustenburg Refiners (Rustenburg) (Pty) Ltd., South Africa. Transferred as Research Metallurgist, Research and Development Department,

with responsibility for improving the performance of a new precious metals recovery plant. Promoted to Assistant Research and Development Manager, April 1975 and Manager, Research and Development Department in May 1976. Responsible for Research and Development projects for an essentially hydrometallurgical operation.

January 1973 - November 1974

Rustenburg Platinum Mines Limited, South Africa (Johannesburg Consolidated Investment Company). Employed as Senior Technical Assistant. Five months shift training on copper-nickel smelter followed by four months training at a concentrator. Took up permanent position at the smelter, commissioning new process in October 1973. Promoted to Acting Plant Metallurgist, April 1974.

September 1968 -

Dorman Long Limited, Teesside (British Steel Corporation). Twelve months' post to gain industrial experience prior to commencing Ph.D. studies. Employed in Research and Development Department, five months in Ironmaking section, seven months in Refractories section.

PUBLICATIONS

Wraith, A.E. & Chalkley, M.E.
'Tuyere Injection for Metal Refining' In 'Advances in Extractive Metallurgy 1977' ed. M.J. Jones (London: IMM, 1977), 27-33.

Chalkley, M.E. & Wraith, A.E.
'Gas Dispersion From a Tuyere in Mercury: Application of Pressure Pulse Analysis'. Trans. Inst. Min. Met., Sec. C, 87, 1978, C266-C271.

Chalkley, M.E. & Wraith, A.E.
'Gas Dispersion at an Annular Tuyere', presented at 'Gas Injection into Liquid Metals' meeting, April 1979, Newcastle upon Tyne.

Chalkley M.E., Doyle B.N., Masters I.M., & Weir D.R.
'Pressure Leaching of New Brunswick Complex Sulphide Concentrate' presented at Canada/EC Seminar, Ottawa, October, 1982.

Chalkley M.E., Masters I.M. & Weir D.R.
'Miniplant Leaching of Zinc Concentrate', presented at Zinc'83, Annual Meeting of Hydromet. Soc., CIM, Edmonton, August 1983.

RPC RESEARCH AND PRODUCTIVITY COUNCIL

PHILIP F. SALIB, B.Sc., M.Sc.Tech., Ph.D., M.Inste., C. Eng.
Head, Fuel Technology

EDUCATION	Ph.D., Sheffield University U.K., Chemical Engineering and Fuel Technology (1978-1982)
	M.Sc.Tech., Sheffield University U.K., Chemical Engineering and Fuel Technology (1976-1978)
	B.Sc. Alexandria University, Egypt, Chemical Engineering (1966-1971)
SPECIALITIES	Coal Combustion and Process Simulation Oil Shale Technology
EXPERIENCE	
1982 - Present	Head, Fuel Technology
1978 - 1982	Coal Combustion Consultant - Coal Research Establishment U.K.
1972 - 1976	Production Engineer-Cociete Industrial Maharrm Press. Alexandria, Egypt. Responsibilities include Raw Material Control and Energy Monitor, Supervision of Manpower and Scheduling of Maintenance.

RELEVANT PROFESSIONAL
EXPERIENCE

While at Sheffied University involved with basic research on spontaneous combustion.

Developed concepts for the exploitation of eastern Canadian oil shales through co-utilization with high sulphur coals.

Designed, constructed and commissioned a circulating fluidized bed combustion system for solid fuels. In 1986 this was the only operating pilot plant in Canada.

Untertook testing of Canadian coals to determine combustion characteristics and sulphur capture obtained with different limestones.

Designed and tested the integrated combustion of coal and oil shale for production of petroleum products, thermal energy for steam raising together with sulphur emission control.

Managed for the New Brunswick Electric Power Commission test program for multi-staged burner for pulverized fuel combustion.

ASSOCIATIONS

Registered Chartered Engineer (UK)
Member Institute of Energy
Member Combustion Institute

PUBLICATIONS

Boorman, R.S., P. Salib, R. Gilders, and D.E. Gemmell, "Fluidised Bed Co-Combustion of Coal and Oil Shales a Route to SO_2 Emission Control", Paper presented at the 3rd Meeting of CIM District 1, Fredericton, Sept. 1982.

Salib, P., and Swithenbank, J. (1981), "Pseduo-Random Stimulus Response of a Rotating Fluidized Bed Combustor", Paper presented at the AIChE 74th. Annual Meeting, New Orleans, Nov. 1981.

Turan, A., Swithenbank, J., Billings, S.A., and Salib, P. (1980), "Pseudo-Random Stimulus Response of Combustion Systems", Journal of Institute of Energy Dec. 1980, p. 181-190.

Salib, P., Swithenbank, J., and Demircan, N. (1980), "The Rotating Fluidized Bed as a High Intensity Combustor", 3rd. Int. Conf. Mech. Power Engng., Monofia University, Egypt, Sept. 1980.

CURRICULUM VITAE

KENNETH RUSSELL ROBSON B.E.Sc., P.Eng.
Chemical Engineer
Department of Applied Sciences

EDUCATION:

M. Eng. University of New Brunswick
Chemical Engineering (in progress)
(1984-Present)

B.E.Sc. University of Western Ontario
Chemical Engineering, 1983

EXPERIENCE:
1985-Present

Chemical Engineer: RPC, Department of Applied Sciences. Coal and Oil Shale Pyrolysis and Combustion in Circulating Fluidized Bed.

Chemical Engineer: RPC, Mineral Development and Processing. Volume Reduction of Low Level Radioactive Waste in a Pyrohydrolysis Reactor.

Characterized wastes produced from the combustion in a CFBC of coal-oil shale mixtures.

ASSOCIATIONS:

Registered Professional Engineer in the Province of New Brunswick

PUBLICATIONS:

P.F. Salib, S.K. Barua, K.R. Robson and R.K. Kissell, "Circulating Fluidized Bed Combustion of Different Ranking Coals with Limestone Sorbents," The Ninth International Conference on Fluidized Bed Combustion, Boston, Massachusetts, May 1987.

C.D. Desjardins, G. Belland, R. Robson and L.P. Buckley, "Demonstration of a Continuous Pyrohydrolysis Waste Volume Reduction System", Canadian Nuclear Society 2nd International Conference on Radioactive Waste Management, Conference Proceedings, Sept. 1986.

APPENDIX B
MICROPROBE AND IMAGE ANALYSIS

MICROPROBE AND IMAGE ANALYSIS

As microprobe and image analysis were outside the scope of the present investigation the work was limited to a single suite of samples. These were taken during Test No. 7 in Task III, which was carried out with Minto coal and Havelock limestone and with dust recycling from the secondary cyclone. The chemical composition of the samples tested are as follows:

Component	Concentration Weight %				
	Cyclone	J-valve	Baghouse	Stand-Pipe	Weepage
Ash	96.78	99.04	91.57	97.78	98.90
Total Carbon	3.47	2.25	8.13	2.92	2.03
Organic Carbon	2.73	0.84	7.26	2.13	0.92
Inorganic Carbon	0.74	1.41	0.87	0.79	1.11
Total Sulphur	9.69	12.20	6.09	10.88	13.12
Sulphation Sulphur	9.9	12.63	6.17	11.14	13.88
Total Calcium	29.86	35.31	16.53	32.45	34.12

The work was of a preliminary nature, however, it led to identification of the most effective particle size range. Future refinement could lead to better understanding of the factor controlling the degree of sulphation of lime particles. It could also be used to expalin the differences in reactivity of different sorbents. Microprobe examination of oil shale could lead to a better understanding of its structure and how it would affect its capacity for sulphur capture.

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MINERALOGICAL REPORT: M/88/002
JOB NUMBER: 08206998/8006
SUBMITTED BY: Department 8-0

MICROPROBE AND IMAGE ANALYSIS STUDY ON FIVE SAMPLES FROM THE FLUIDIZED BED COMBUSTOR CIRCUIT AT RPC

Prepared by:

Lech Lewczuk
Petrologist

LL/mb/MINRE1

1.0 INTRODUCTION

Samples of the material from the following five points of the Fluidized Bed Combustor at RPC were analyzed:

1. J. Valve
2. Standpipe
3. Weepage
4. Secondary Cyclone
5. Baghouse

The object of the analysis was to evaluate the sulphur capture by the limestone particles during the combustion of the limestone/coal blend and to study the influence of the limestone grain size on the process.

2.0 PRELIMINARY EXAMINATION OF THE SAMPLES

Listed below are the results of the preliminary optical and electron microprobe examination of the samples received.

2.1 Samples from the standpipe, J. valve and weepage are characterized by a wide range (from minus 5 microns to plus 1000 microns) of grain sizes.

2.2 Samples from the secondary cyclone and baghouse are much finer grained, with grain size ranging from minus 5 microns to plus 106 microns.

2.3 Grains of the original limestone are completely converted (oxidized) to lime (CaO) in each sample. Lime, in turn, is affected by sulphation to a variable degree.

2.4 Sulphation (penetration of lime grains by SO_2 released from coal) leads to the formation of a calcium-oxygen-sulphur (Ca-O-S) compound. Most of the lime grains are replaced by this compound, but accessory amounts of lime free of sulphur were also observed.

2.5 Lime (CaO) free of sulphur occurs either as very fine (minus 100 microns) individual grains, or it forms portions of grains (usually their interiors) that are affected by sulphation.

2.6 Quantitative, standardless analysis of the Ca-O-S compound shows that it contains sulphur in highly variable amounts that range from below 1% to 24%.

2.7 Figure 1 shows an analytical traverse across a typical sulphated grain of lime. The figure indicates that the rims of the grain are composed of a sulphur-rich Ca-O-S compound, while the core of the grain contains less than 5% sulphur. The distribution of

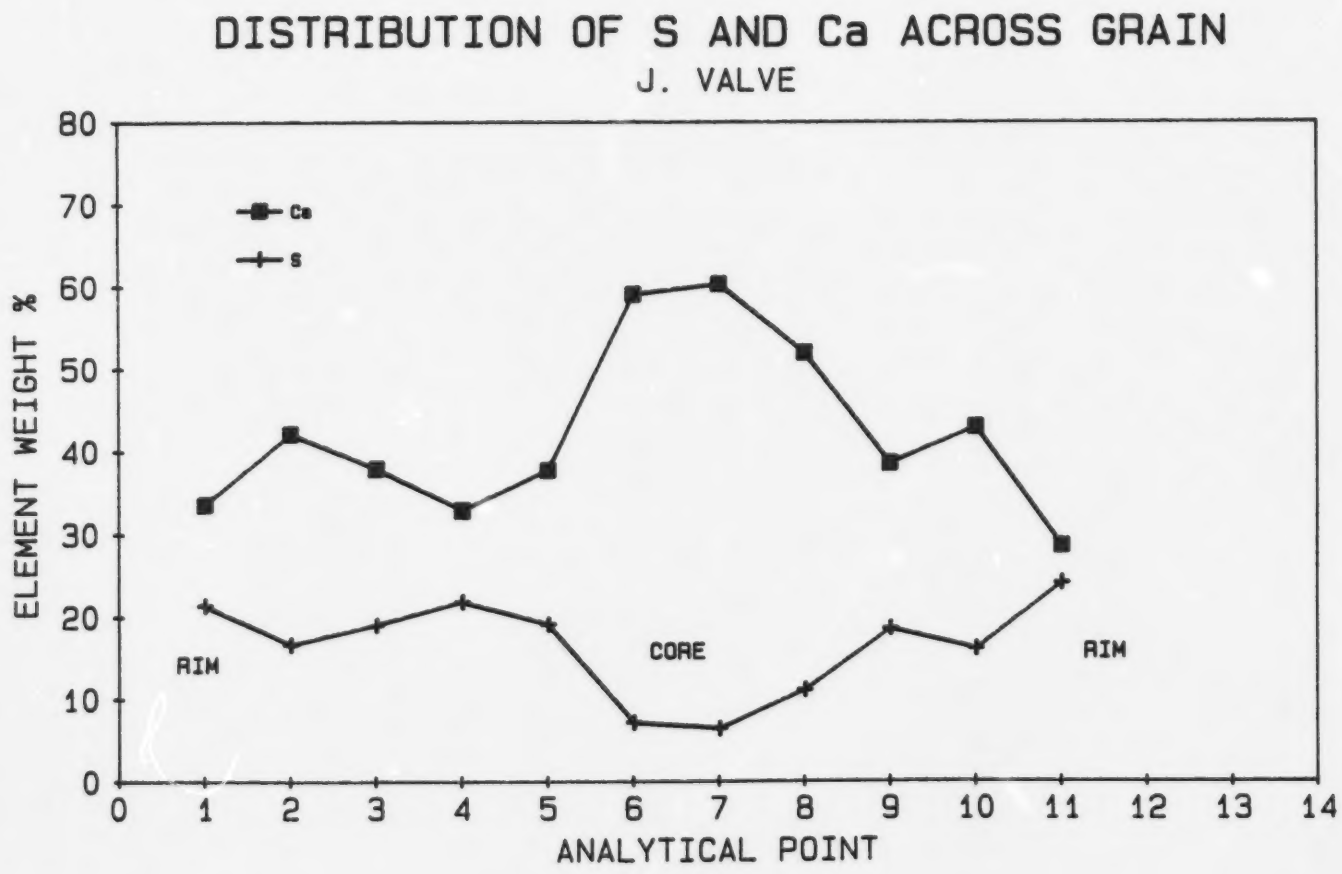

DISTRIBUTION OF S AND Ca ACROSS GRAIN
J. VALVE
ELEMENT WEIGHT %
ANALYTICAL POINT
Ca
S
RIM
CORE
RIM
0
10
20
30
40
50
60
70
80
0
1
2
3
4
5
6
7
8
9
10
11
12
13
14

FIGURE 1

sulphur and calcium in sulphated lime grains is usually less symmetrical than in Figure 1 (compare Mineralogical Report M/87/040 from February 10, 1988).

2.8 Figure 2 displays a compositional image of a typical grain of lime. The grain consists of an intensely fractured and porous interior (A), a sulphur rich rim (B) and an incompletely developed film of ash (C) on its surface. The interior of the particle contains 0.5% - 5% sulphur while the concentration of this element in the rim ranges from 15% - 20%. This is shown in Figure 2A, where the density of white dots reflects the concentration of sulphur.

2.9 Examination of the grains from each sample suggest that the borders between zones A, B and C are clearly visible in the compositional image of the grains, if all three zones occur in a given grain. This is because there are very narrow "intermediate" zones between cores (A) and rims (B) containing 5-15% sulphur. Therefore, as far as the composition of a given grain is concerned, this "intermediate" zone may be neglected. The border between the fly ash (C) and zone B is well defined because of the high content of iron in the fly ash.

2.10 It was not possible to define a compositional border between "pure" CaO and Ca-O-S compound characterized by low sulphur content values (below 5%) in a measurable image of the grains.

2.11 Quartz has been observed in all samples examined. It originates from the limestone and coal and forms either free grains or is intergrown with sulphated and non-sulphated lime, within composite grains.

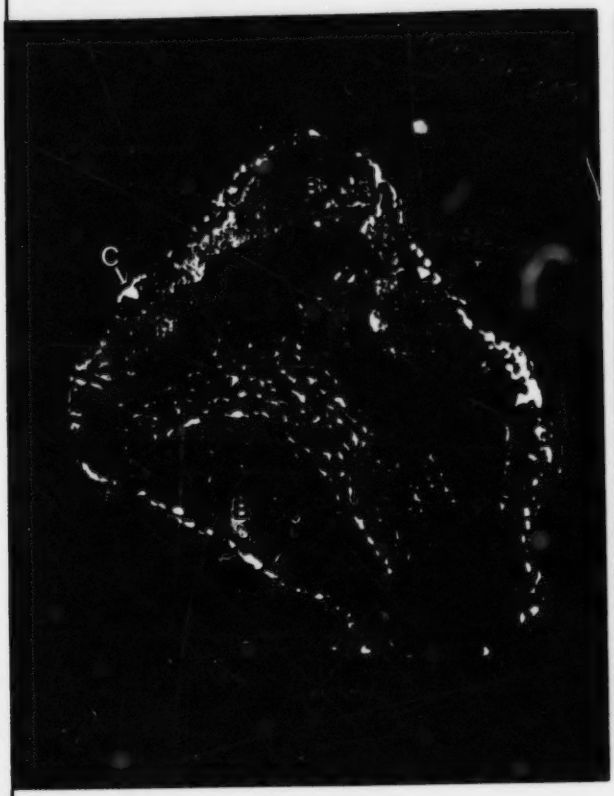


FIGURE 2
Sulphated grain of lime
A: low sulphur Ca-O-S
B: high sulphur Ca-O-S
C: fly ash

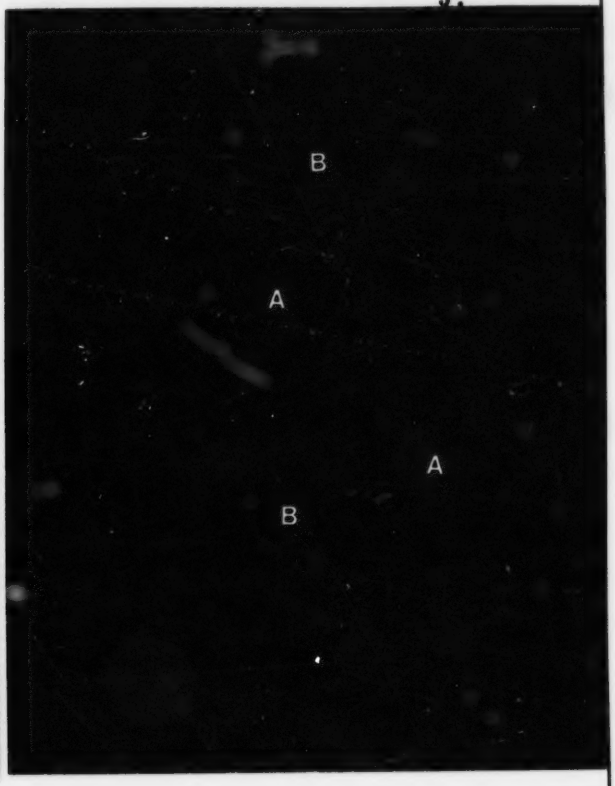


FIGURE 2A
Map of sulphur in grain from Figure 2.
A: low sulphur Ca-O-S
B: high sulphur Ca-O-S

Scale bar = 10 microns

3.0 MEASURED PARAMETERS OF THE SAMPLES

Based on the preliminary examination of the samples, (2.1-2.11), it was assumed that the content of a "high sulphur" compound in which the concentration of sulphur exceeds 15% in a given lime grain may serve as a measure of the efficiency of this grain with regard to the sulphur capture. In other words, the higher the areal ratio of zone B to zone A (which may contain 0%-5% of sulphur) the more efficient that sulphur capture has been within the grain.

Eventually, the following parameters characterizing each sample were measured:

3.1 Relative quantities (area) of each of the following components:

1. High sulphur ($\geq$15% S) Ca-O-S compound
2. Low sulphur ($\leq$5% S) Ca-O-S compound and CaO
3. Quartz and trace amounts of other silicates
4. Fly ash.

3.2 Size distribution of the grains that contain high sulphur Ca-O-S and low sulphur Ca-O-S compound (grains composed of fly ash and quartz only were not measured).

3.3 In addition to the above, the grains were classified based on the content of the high sulphur Ca-O-S compound.

The measurements were performed on the polished grain mounts using an Electron Probe Micro Analyzer interfaced with a KONTRON Image Analyzer. One grain mount was produced from each sieve fraction of each sample except for the baghouse since the sample of the latter was not sized. Upon completion of the analysis, the data generated was combined to represent a whole sample. The sample from the baghouse was not sized due to the relatively narrow range of sizes of the particles present in the sample.

4.0 RESULTS

4.1 Sieve Analysis

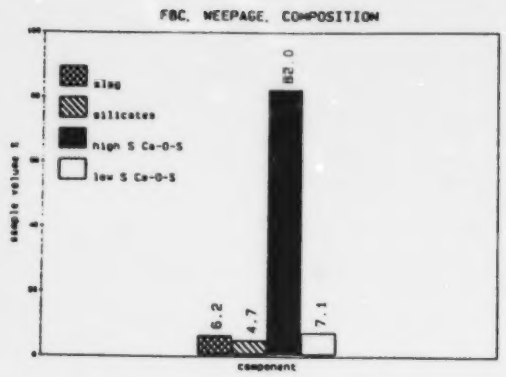
The results of the sieve analysis of the samples are reported in Table 1.

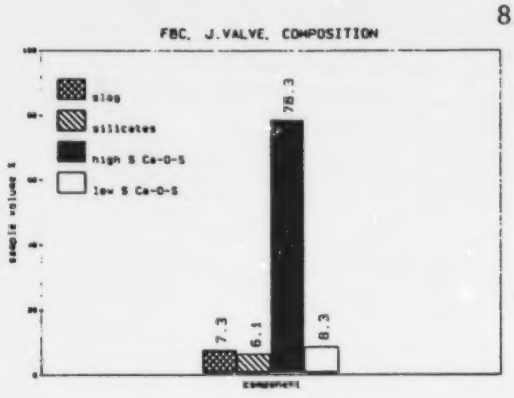
4.2 Composition of the Samples

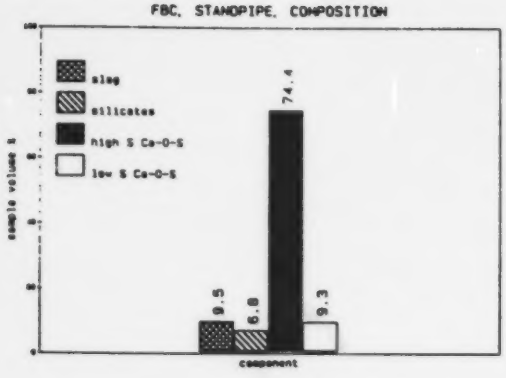
The composition of the samples is shown in Figure 3. The graphs illustrate areal (or volumetric) proportions of the components in each sample. The proportions are expressed as percentages of the total area of the grains measured in a given sample.

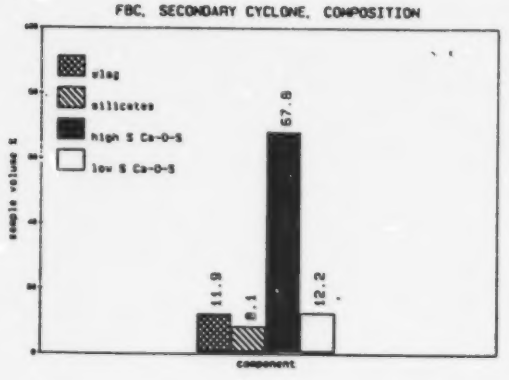
Weight percentages of the components in the samples were not calculated due to the nonstoichiometric composition of the Ca-O-S compounds, lack of data on the density of these compounds and their variable, often very high porosity. It should be noted that high porosity, especially in the low sulphur Ca-O-S compound, could have biased the data on the content of this compound in five coarse (plus 106 microns) fractions of the samples. In these fractions, a low magnification was used in order to measure the size of the grains. Consequently, the pores were not identified as empty fields, but measured as if they were a low sulphur Ca-O-S compound.

The quantitative data reported in Figure 3 may serve as a basis for the material balance calculations but such calculations would require more data on the average composition of the Ca-O-S compounds and their physical properties.


FBC. WEEPAGE. COMPOSITION
slag
silicates
high S Ca-O-S
low S Ca-O-S
6.2
4.7
82.0
7.1
component


FBC. J.VALVE. COMPOSITION
slag
silicates
high S Ca-O-S
low S Ca-O-S
7.3
6.1
78.3
8.3


FBC. STANDPIPE. COMPOSITION
slag
silicates
high S Ca-O-S
low S Ca-O-S
9.5
6.8
74.4
9.3
component


FBC. SECONDARY CYCLONE. COMPOSITION
slag
silicates
high S Ca-O-S
low S Ca-O-S
11.9
8.1
67.8
12.2
component

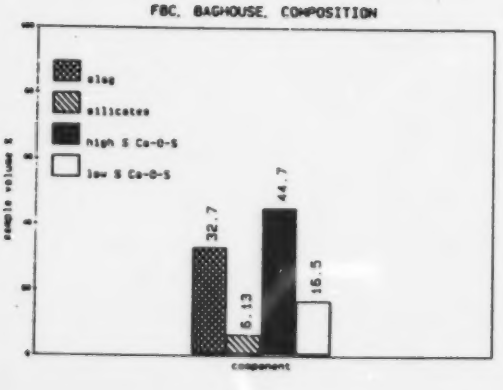

FBC. BAGHOUSE. COMPOSITION
slag
silicates
high S Ca-O-S
low S Ca-O-S
32.7
6.13
44.7
16.5
component

FIGURE 3

Table 1. Size Analyses of Fluidized Bed Combustor Samples

Sample	Mesh	+16	+42	+150	-150
	Microns	+1000	+355	+106	-106
Standpipe		3.3	13.2	15.2	68.3
J.V.		12.1	51.9	32.2	3.8
Weepage			90.2	9.8	
Sec. cyclone			.4	5.1	94.4

Baghouse: not sized.

4.3 Grain Size Distribution

Grain size distribution for each sample is reported in the form of the graphs in Figures 4-8.

Each graph contains a pair of bars ascribed to a given grain size class. The open bar represents the grain size distribution of all grains in the sample containing calcium. The solid bar gives the distribution of grains, in which more than 95 area (volume) % is occupied by the high sulphur Ca-O-S compound.

The grains, in which more than 95 area (volume) % is occupied by a high sulphur Ca-O-S compound were identified as completely sulphated grains of lime. It should be emphasized that the term "completely sulphated" defines only the highest level of saturation with sulphur which may be achieved by the grains of lime in certain conditions of the coal/lime combustion.

A total of twelve grain size classes were established (in microns):

+0; +5; +8; +24; +41; +70; +120; +204; +347; +589; +1000.

4.4 Classification of Grains Based on the Content of High Sulphur Ca-O-S Compound. (Sulphur Capture Classes of Grains.)

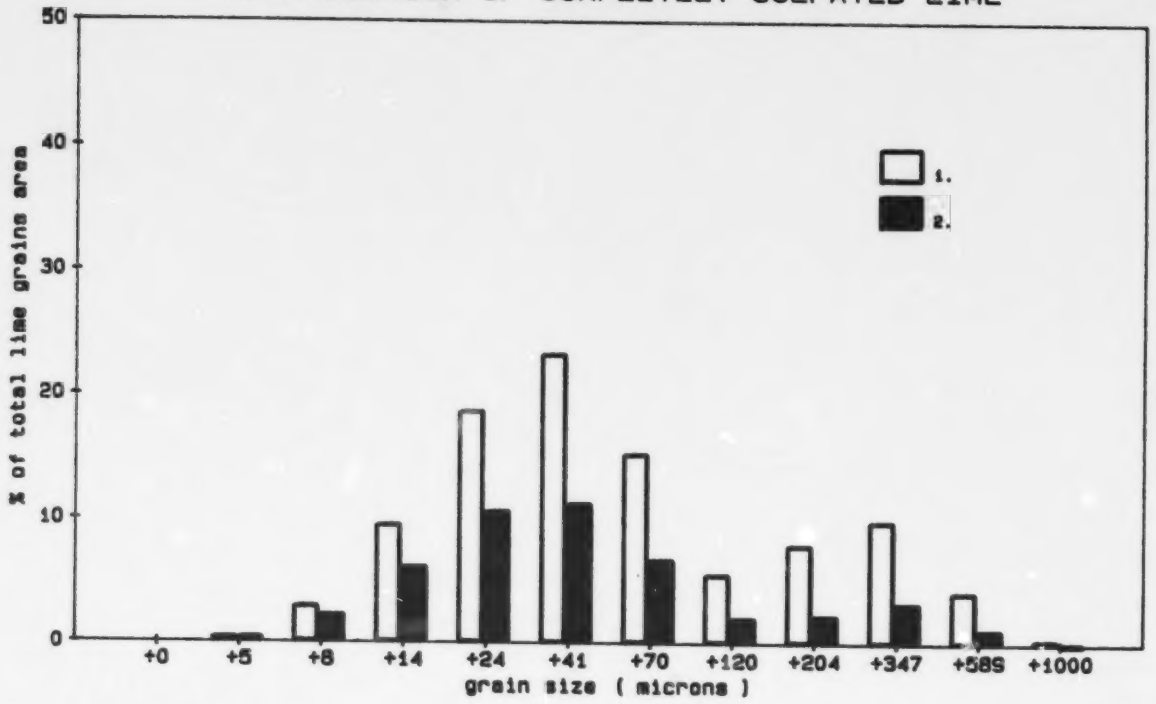
The data is reported in Figures 9-19. The figures present the data for each of the twelve classes in all samples. When the total area of a grain size class was less than 2% of the area of the total sample the data was not considered. In the graphs, individual curves represent individual samples.

The data points on the curves represent percentages of the total area (volume) of the grains, in that size class, that occur in one of the ten classes of sulphur content. Below is the list of these classes:

+10	+55
+15	+65
+25	+75
+35	+85
+45	+95

Class +10 defines a grain, in which at least 10% but less than 15% of the area (volume) is occupied by the high sulphur Ca-O-S compound. Grains in the +95 class may be considered as "completely sulphated" grains (compare also 4.3). For example, the grain shown in Figure 2 is of +50 class, because about 50% of its area (volume) is occupied by high sulphur Ca-O-S compound (zone B).

FBC. STANDPIPE. COMBINED FRACTIONS
SIZE DISTRIBUTION OF LIME GRAINS
DISTRIBUTION OF COMPLETELY SULFATED LIME

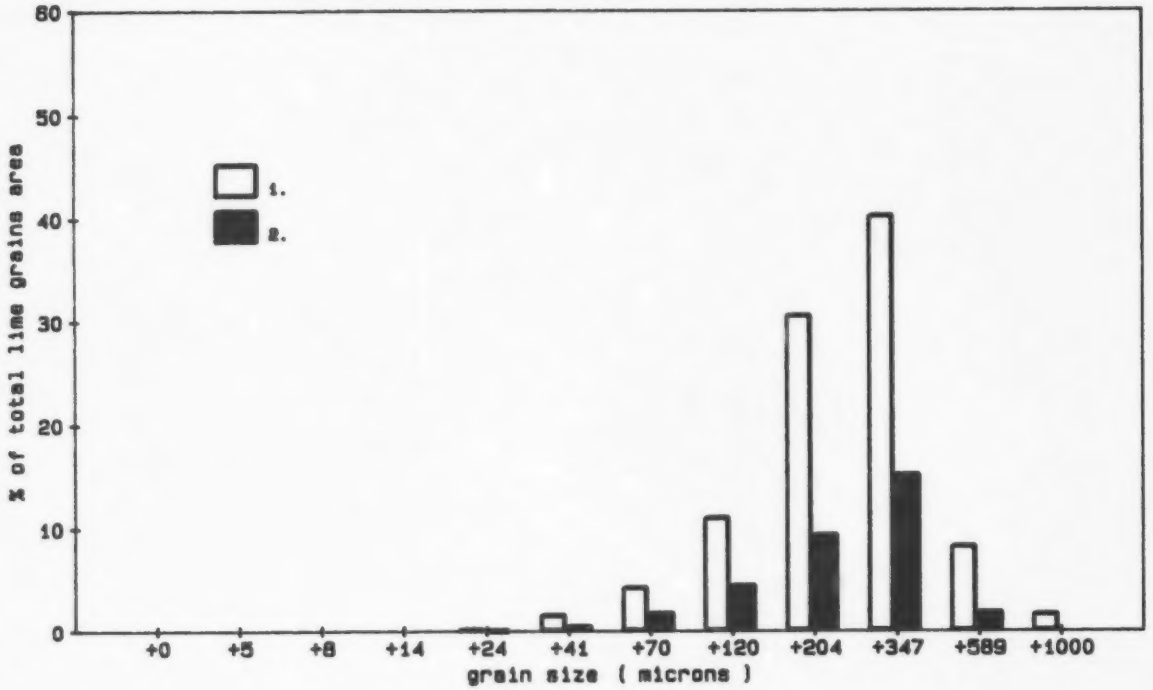


LEGEND

1. ALL LIME GRAINS IN SAMPLE
2. COMPLETELY SULFATED LIME GRAINS

FIGURE 4

FBC. J.VALVE. COMBINED FRACTIONS
SIZE DISTRIBUTION OF LIME GRAINS
DISTRIBUTION OF COMPLETELY SULFATED LIME

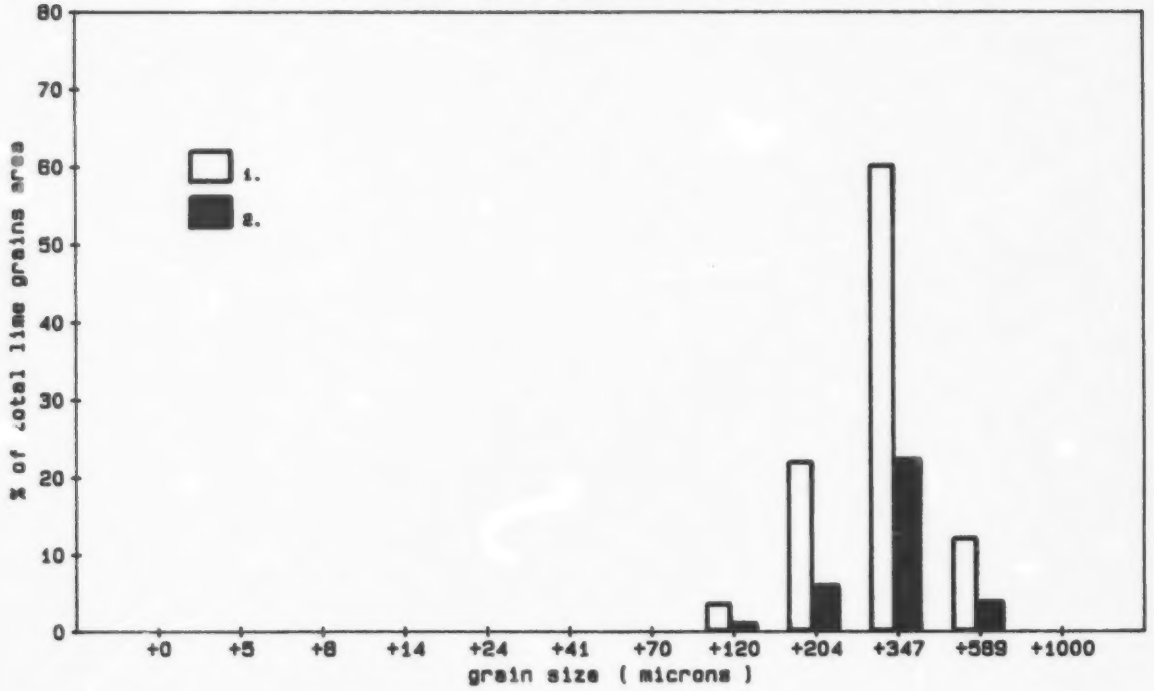


LEGEND

1. ALL LIME GRAINS IN SAMPLE
2. COMPLETELY SULFATED LIME GRAINS

FIGURE 5

FBC, WEEPAGE, COMBINED FRACTIONS
SIZE DISTRIBUTION OF LIME GRAINS
DISTRIBUTION OF COMPLETELY SULFATED LIME

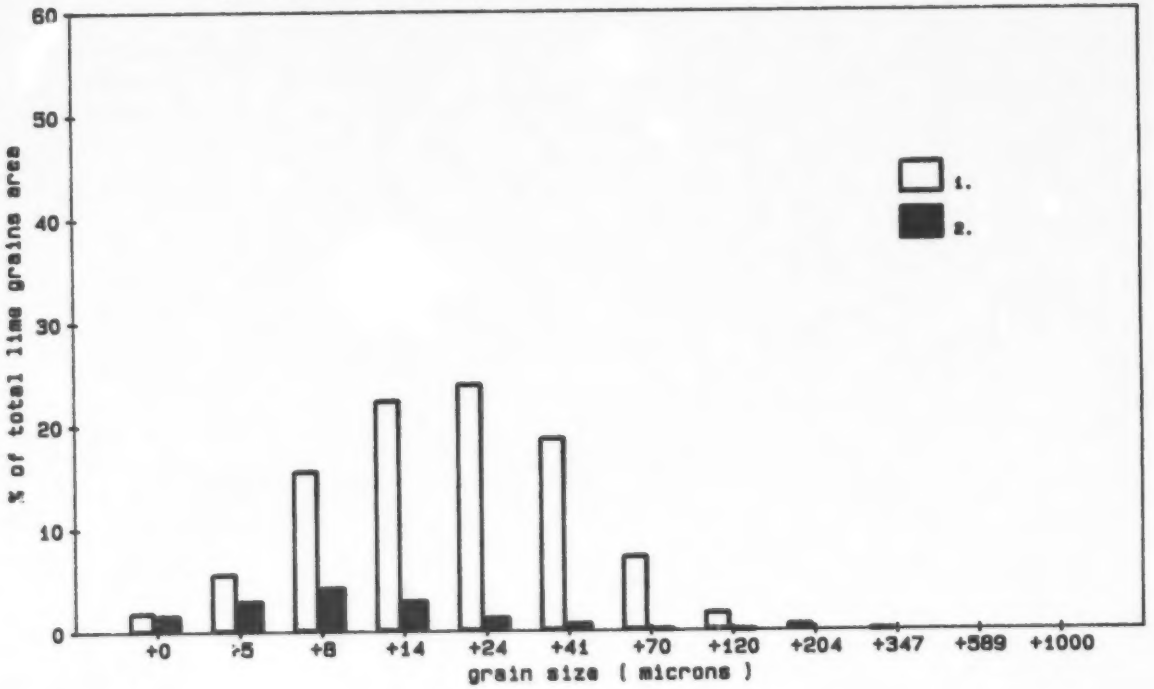


LEGEND

1. ALL LIME GRAINS IN SAMPLE
2. COMPLETELY SULFATED LIME GRAINS

FIGURE 6

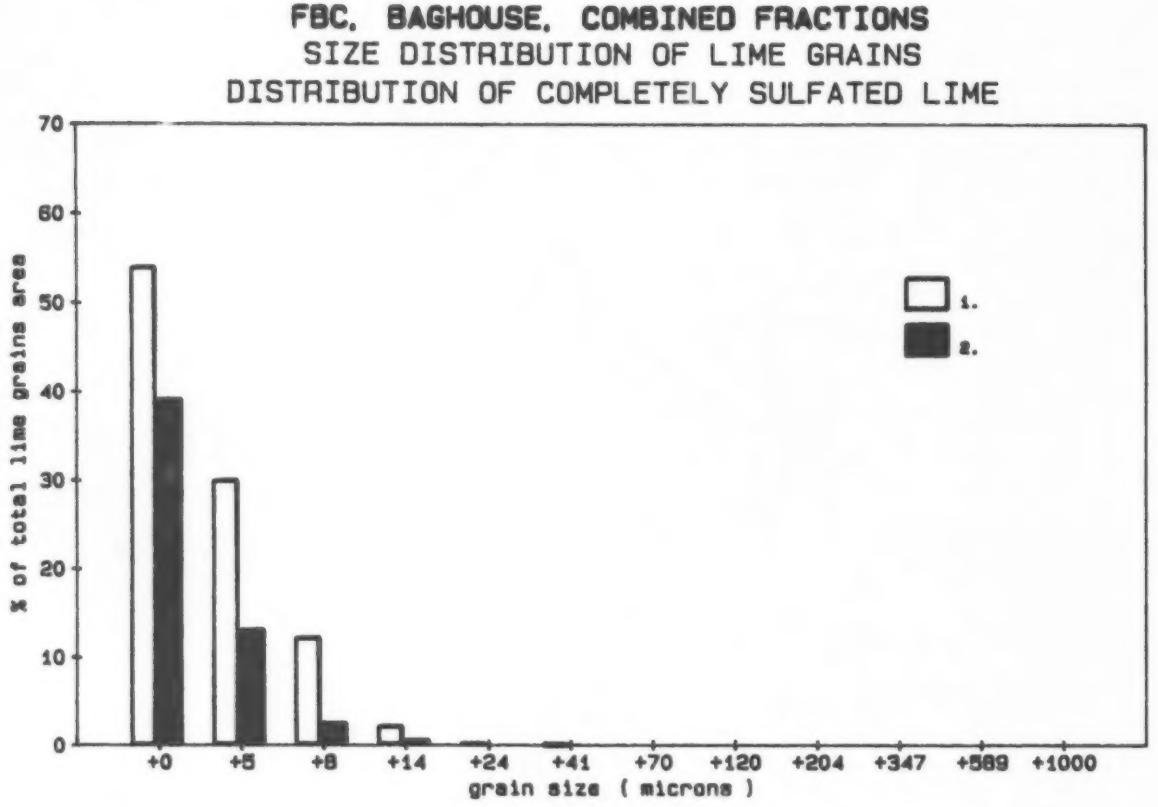
FBC. SECONDARY CYCLONE. COMB.FRACTIONS
SIZE DISTRIBUTION OF LIME GRAINS
DISTRIBUTION OF COMPLETELY SULFATED LIME



LEGEND

1. ALL LIME GRAINS IN SAMPLE
2. COMPLETELY SULFATED LIME GRAINS

FIGURE 7

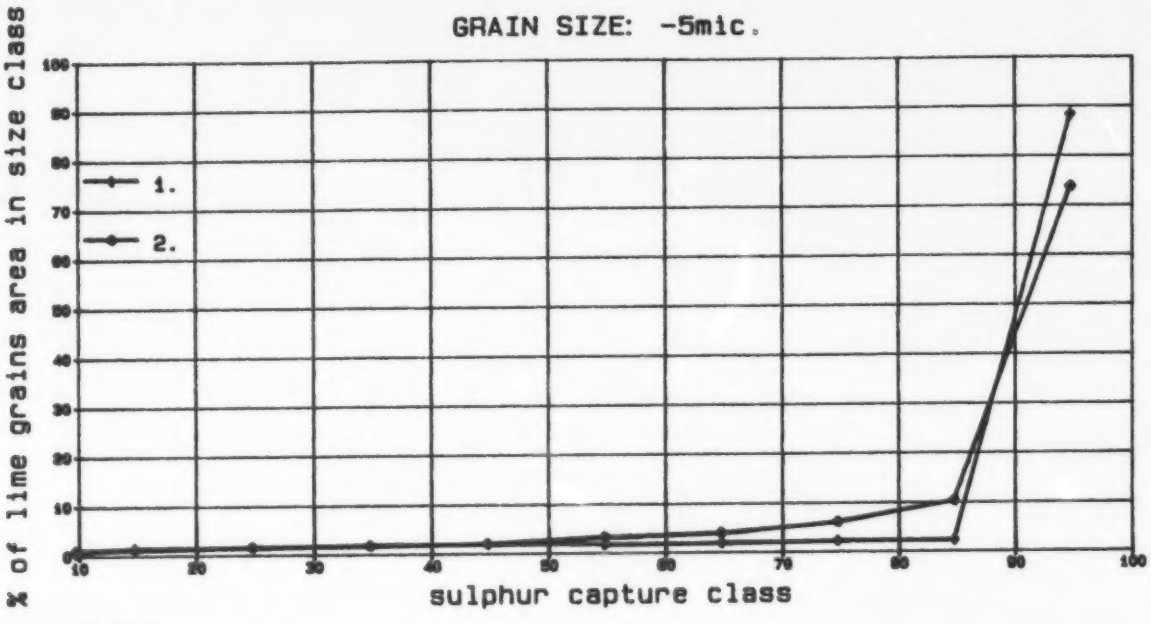


LEGEND

1. ALL LIME GRAINS IN SAMPLE
2. COMPLETELY SULFATED LIME GRAINS

FIGURE 8

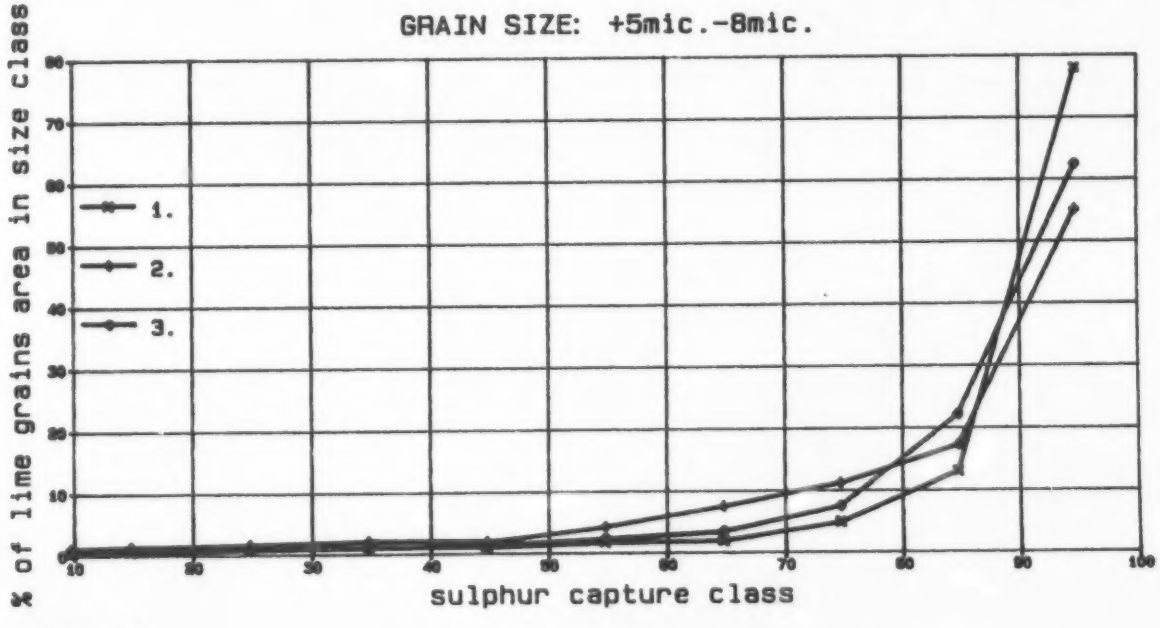
FBC. SULPHUR CAPTURE



SAMPLE

1.SECONDARY CYCLONE
2.BAGHOUSE

FIGURE 9



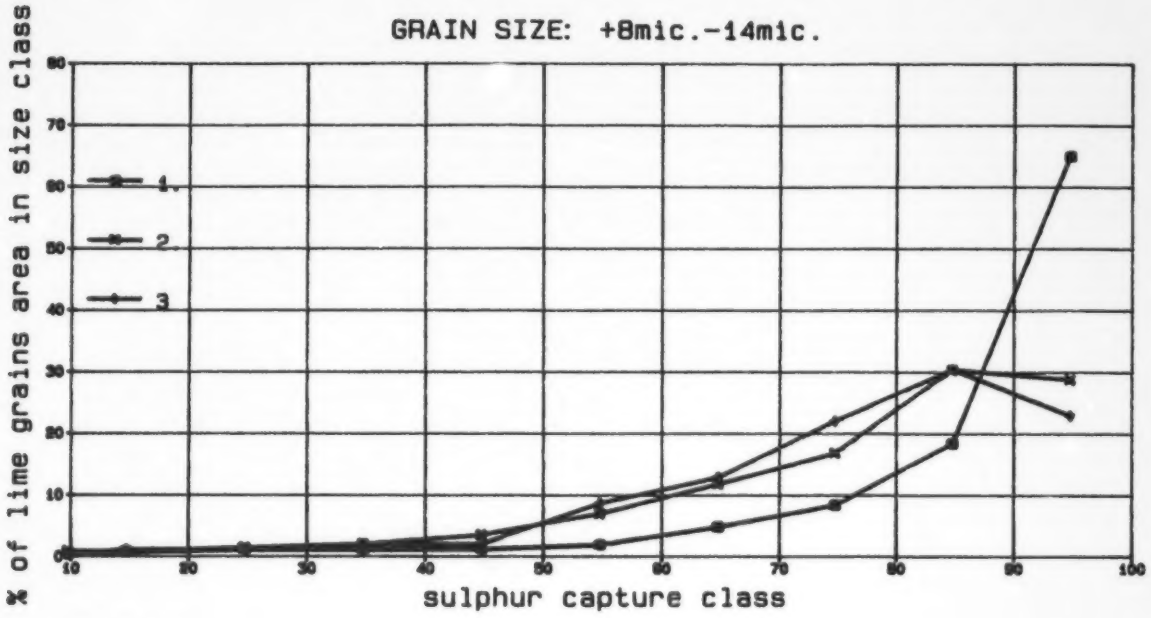
SAMPLE

1.STANDPIPE
2.SECONDARY CYCLONE
3.BAGHOUSE

FIGURE 10

FBC. SULPHUR CAPTURE

GRAIN SIZE: +8mic.-14mic.

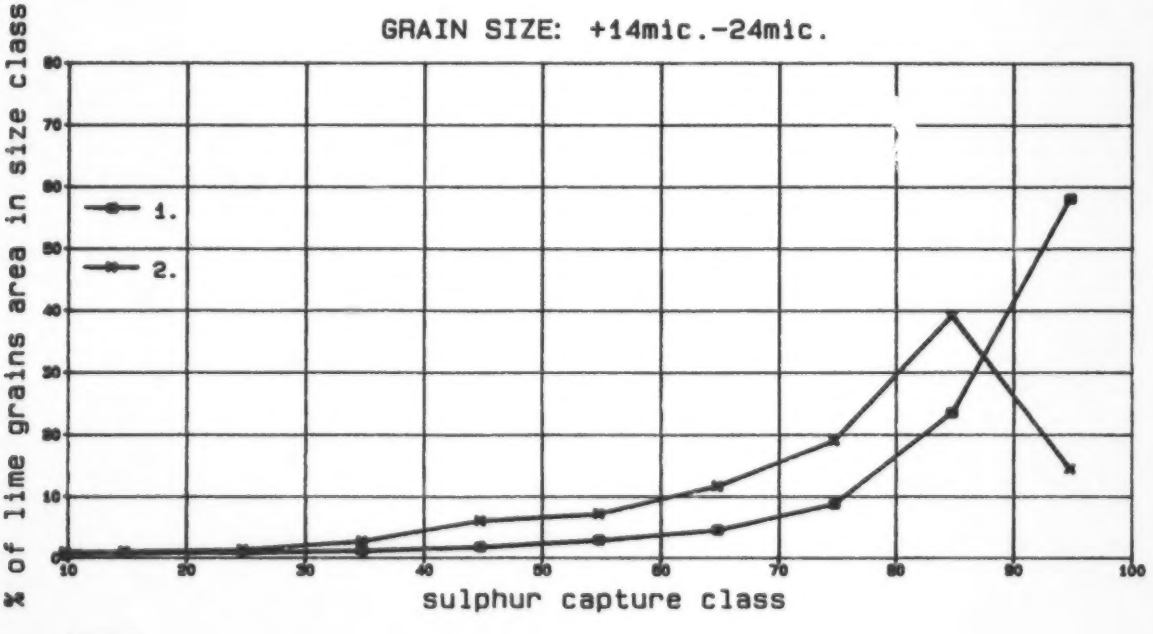


SAMPLE

1.STANDPIPE
2.SECONDARY CYCLONE
3.BAGHOUSE

FIGURE 11

GRAIN SIZE: +14mic.-24mic.



SAMPLE

1.STANDPIPE
2.SECONDARY CYCLONE

FIGURE 12

FBC, SULPHUR CAPTURE

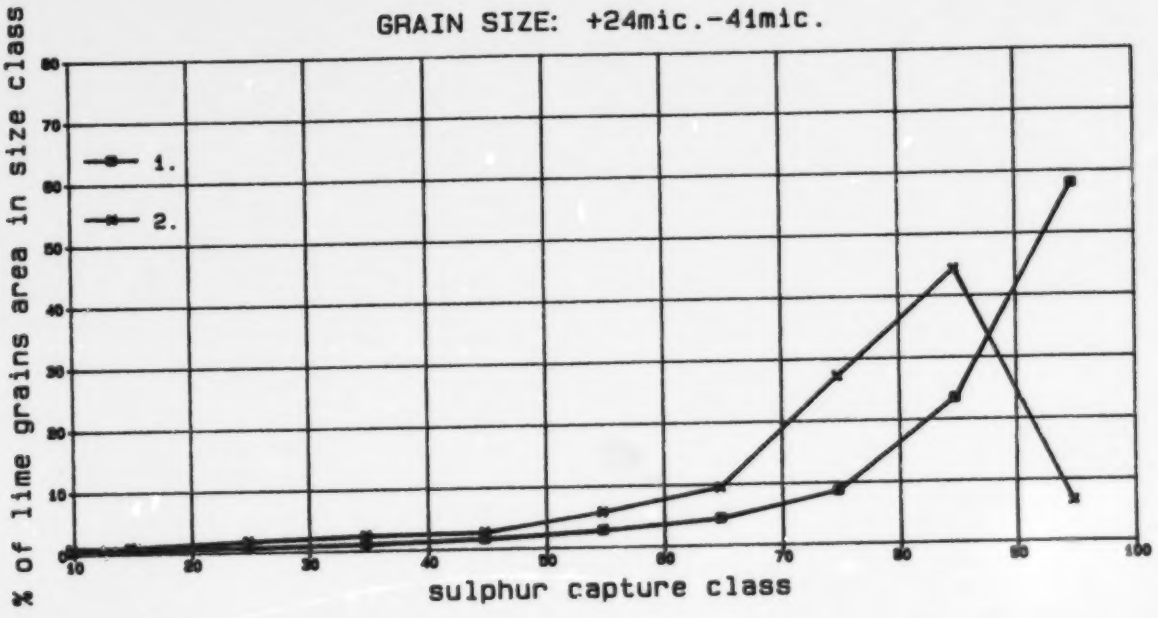


FIGURE 13

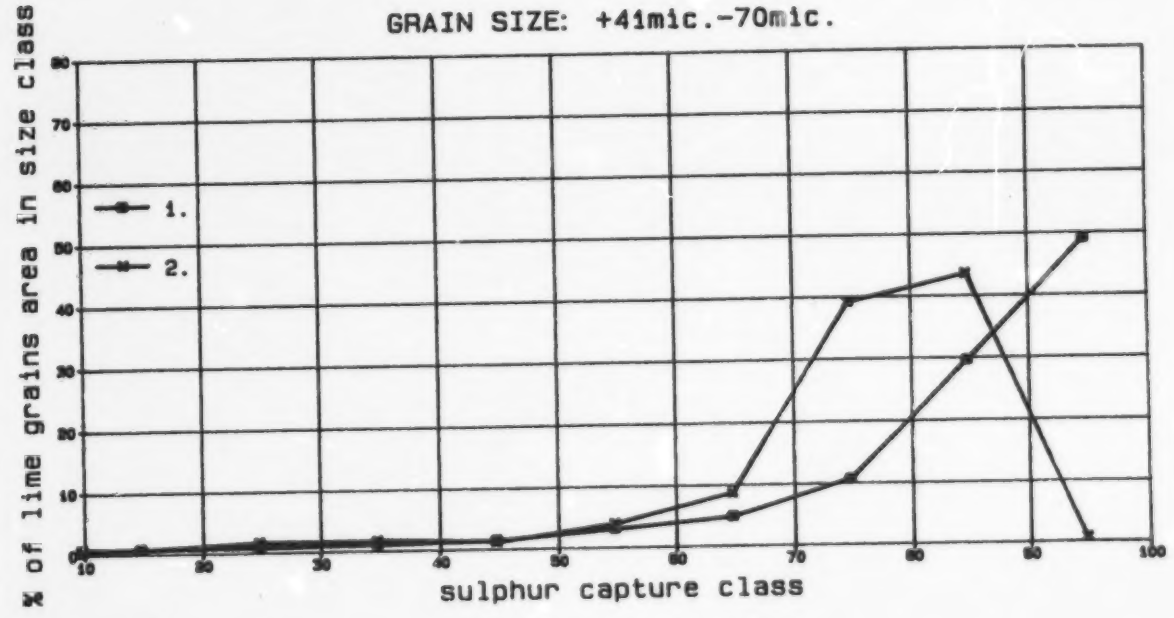
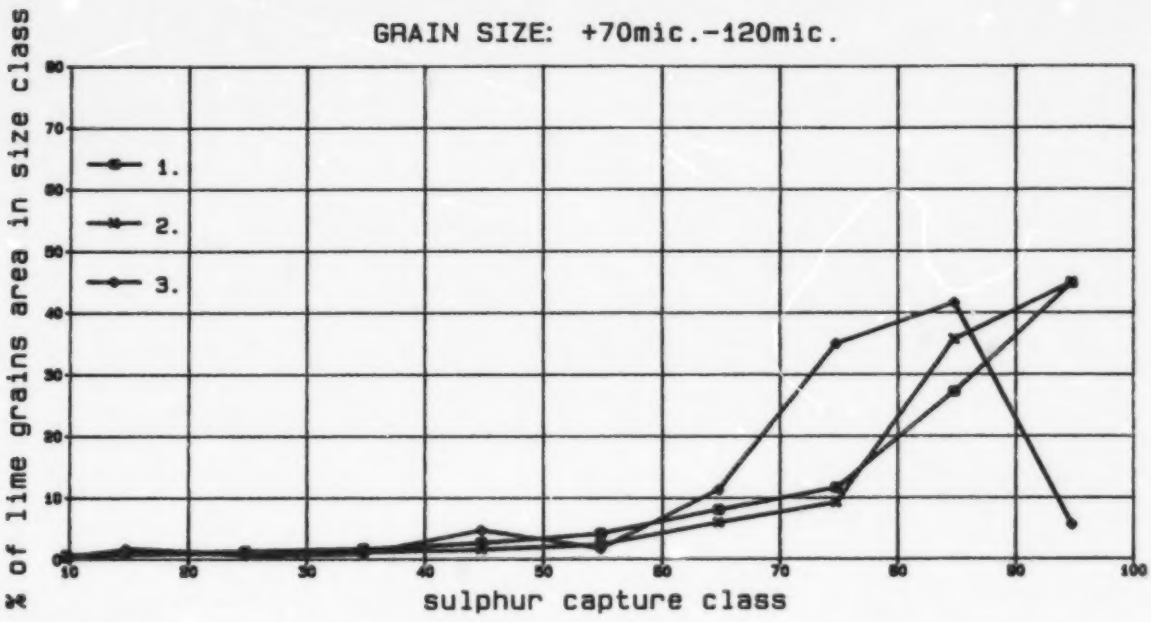


FIGURE 14

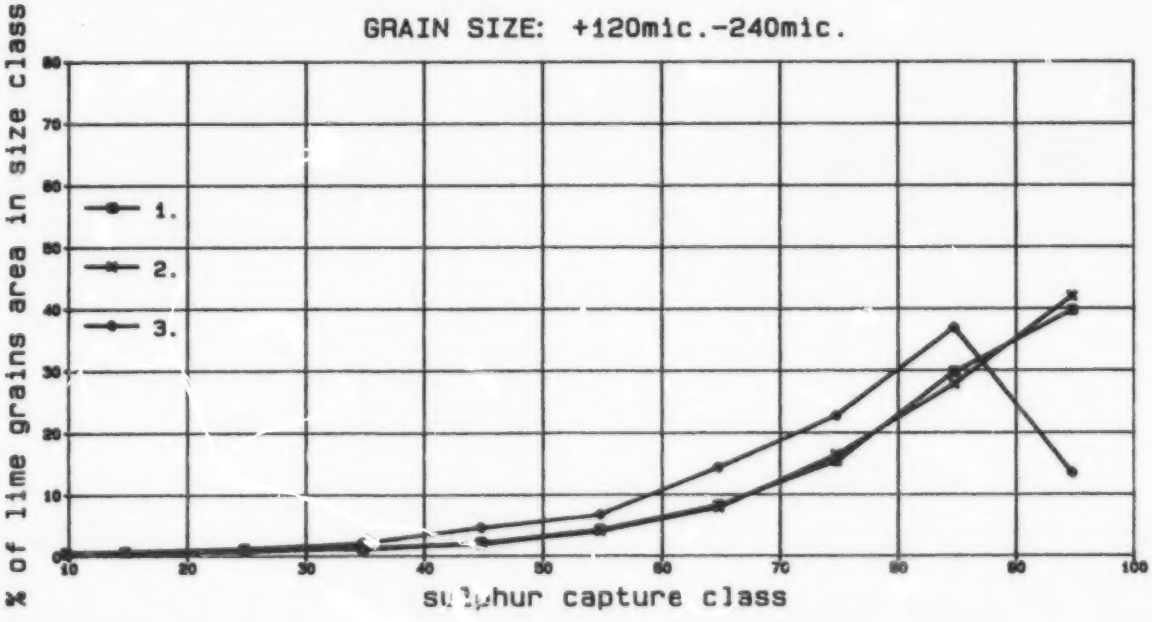
FBC. SULPHUR CAPTURE



SAMPLE

1.J.VALVE
2.STANDPIPE
3.SECONDARY CYCLONE

FIGURE 15



SAMPLE

1.STANDPIPE
2. J.VALVE
3.SECONDARY CYCLONE

FIGURE 16

FBC, SULPHUR CAPTURE

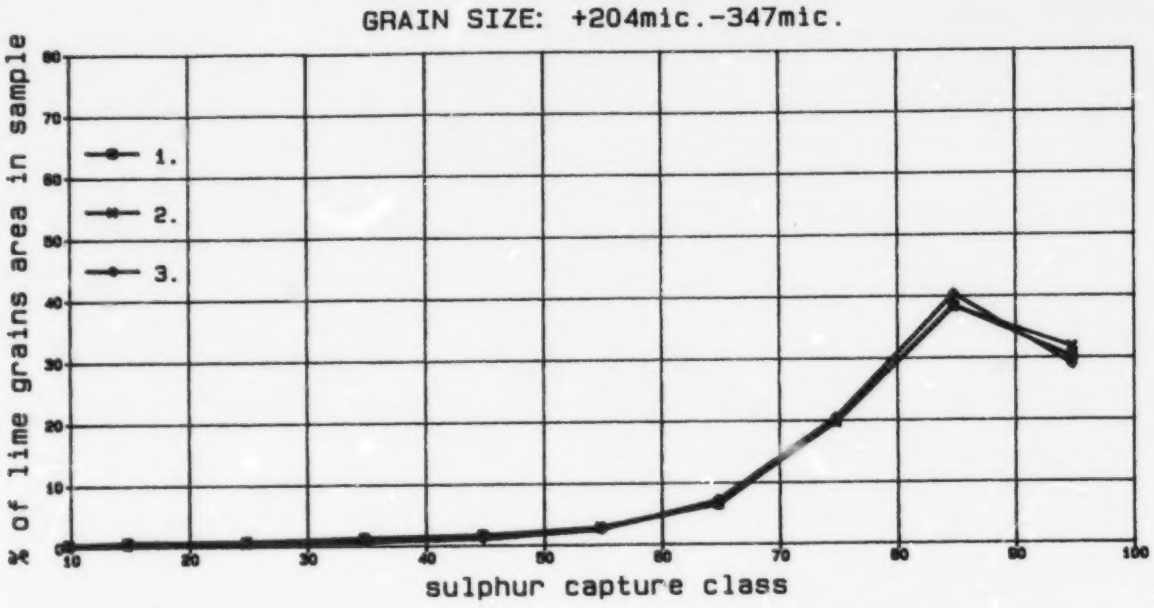


FIGURE 17

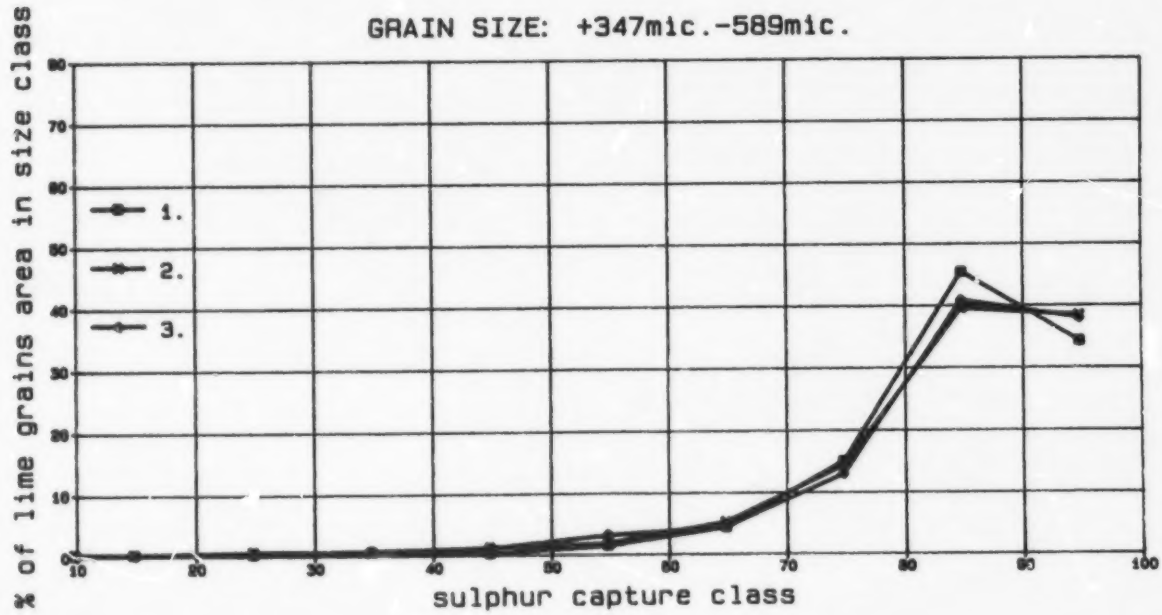
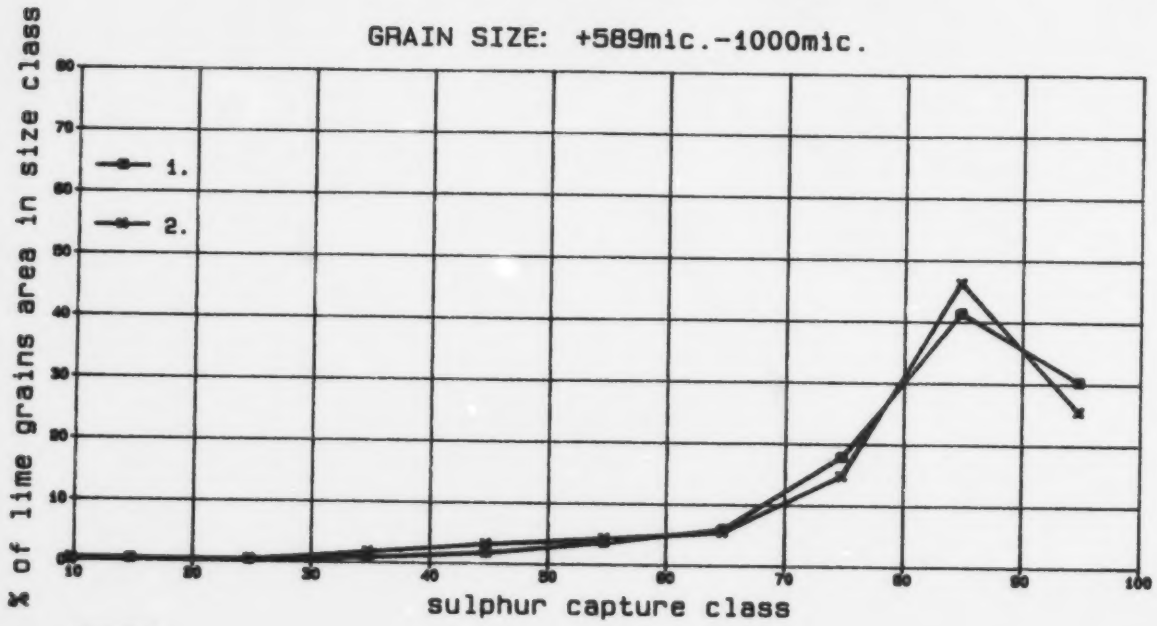


FIGURE 18

FBC. SULPHUR CAPTURE



SAMPLE

1.STANDPIPE
2. J.VALVE

FIGURE 19

5.0 MORPHOLOGY OF LIME AND FLY ASH GRAINS

Stereo microscope examination of the samples revealed as follows:

5.1 A thin (usually not thicker than 1/10 of the diameter of a grain) film of fly ash covers numerous sulphated grains larger than 100 microns, in the samples of the J valve, standpipe and weepage.

5.2 Only a few grains finer than 50 microns were observed to be covered with fly ash.

5.3 Fly ash frequently forms thin layers on the grains of quartz and other silicates.

5.4 Fly ash occurs also as "free" (not combined with any other constituent) grains of diversified morphology.

5.5 The most common form of the fly ash grains is as highly porous, irregular aggregates. The size of the grains does not exceed 30-40 microns. Spheres (cenospheres and solids) often as large as 1000 microns, were also observed (Figure 20, grains A).

5.6 Most of the grains of lime larger than about 500 microns are rounded and subrounded regardless of the thickness of the layer of fly ash developed on their surface. Well rounded grains (Figure 20, grains B), barely covered or not covered with fly ash were found in the samples from J valve, standpipe and weepage. It was noticed that finer grains tend to be more angular than the larger ones.

5.7 Disintegration of large, +500 micron grains in the samples from the standpipe, J valve and weepage was observed. The phenomenon is always associated with swelling and cracking of the grains (Figure 20, grains B and C). The cores of these grains contain

FIGURE 20.

Grains A: cenospheres of fly ash from standpipe.

Grains B: lime grains noncovered and poorly covered with fly ash (dark grey). Note swelling of grains and their cracking.

Grains C: Disintegration of lime grains with thick layer of fly ash (black) on their surface. Note that some of the grains are not cracked.

Scale bar = 3.0 mm

either low sulphur Ca-O-S compound or CaO, both extremely porous. Regardless of the causes (degasification?, thermal shock?, attrition?), the phenomenon leads to the formation of very fine grains of either sulphated or only slightly sulphated lime, and "free" grains of fly ash. Both the low sulphur lime and fly ash grains that are derived from the disintegration of larger grains in the fluidized bed contribute to the high content of low sulphur Ca-O-S compound and fly ash in the samples from the secondary cyclone and the baghouse. Nevertheless, most of the fly ash in the baghouse results from coal combustion, as indicated by the shape of its grains.

6.0 DISCUSSION

Based on the data generated, the following has been concluded.

6.1 For those particular samples all the grains of limestone calcined to lime (CaO) as indicated by the absence of calcium carbonate ($CaCO_3$) in the samples.

6.2 The lime absorbs SO_2 released during the coal combustion. The process leads to the formation of a calcium-oxygen-sulphur compound. The compound contains variable quantities of sulphur ranging from below 1% to 24%.

6.3 Two major compositional types of the Ca-O-S compound may be distinguished. They are low sulphur Ca-O-S containing up to 5% sulphur and high sulphur Ca-O-S with 15%-20% sulphur.

6.4 Both the low sulphur and high sulphur Ca-O-S compounds are the major constituents of the sulphated lime grain. The volumetric ratio of these constituents in a grain is a measure of the efficiency of this grain with regard to the sulphur capture.

6.5 Lime grains converted completely to high sulphur Ca-O-S compound may be considered saturated with sulphur. As such, they are inactive (incapable of capturing more sulphur) under the combustion conditions existing during the sampling campaign.

6.6 The coarser samples are (relatively) richer in the high Ca-O-S compound (Figures 4-8, compare Figure 3). This is because the finest grains may be removed to the secondary cyclone overflow and eventually the baghouse before they are saturated with sulphur.

6.7 In the process of combustion, the grains of lime disintegrate regardless of whether they are completely sulphated or not. As a result, finer grains of lime and sulphated lime are produced and a portion of these grains passes over to the baghouse. It is also believed that the grains of (oxidized) limestone whose size ranges from 0 to about 40 microns may not stay long enough in the system to absorb any sulphur and that they contribute to the high content of non-sulphated lime particles in the baghouse. In order to prevent the latter, the limestone feed should not contain grains finer than about 40 microns. In other words, the whole baghouse material should be derived from the disintegration of the lime grains during combustion.

6.8 The degree of sulphation of the grains exposed to the combustion gases is invertly proportional to their size (Figures 4-8).

6.9 The lime grains affected by sulphation from the J valve, standpipe, weepage and secondary cyclone underflow have such a composition that at least 35% of their volume is occupied by high sulphur Ca-O-S compound regardless of their size (Figures 9-19). This is indicated by the close to zero quantities of the +10, +15, +25 and +35 classes of sulphur capture in the individual size classes. Such grains are potentially "active", or able to capture more sulphur if not removed to the baghouse.

6.10 Even the largest, (+589 microns) grains of lime may become completely replaced by high sulphur Ca-O-S compound as shown by the quantities of the class +95 of sulphur capture (Figure 19). This requires that these grains remain in the system for a sufficiently long (undefined) period of time.

6.11 It is not known whether all lime grains, if kept in the system for a sufficiently long period of time, would convert to the high sulphur Ca-O-S compound. It is possible that the development

of a layer of fly ash on the grains coarser than 50 microns, may slow down or completely halt the sulphation process. Attrition or disintegration of the particles would be required to reactivate them.

6.12 There is strong evidence that the process of disintegration is rather violent (Figure 20) and results in the formation of fractions finer than 40 microns, which quickly enter the baghouse. It is also possible that the process of sulphation may be influenced by such factors as the chemical composition and physical properties (microstructure) of the limestone grains.

6.13 The disintegration of partly sulphated grains may produce fine grains of both active (partly non-sulphated) grains and inactive grains (highly sulphated). However, these finer grain sizes may quickly report to the baghouse and leave the system.

6.14 The grains from the secondary cyclone whose size range from +14 microns to -120 microns are absent (Figure 14) or almost absent (Figures 13, 15, 16) in the +95 class of sulphur capture. On the other hand, at the time of sampling, between 45 and 60 percent of the grains of this size in the J valve and J standpipe samples are completely sulphated.

7.0 CONCLUSIONS

Based on the data generated, the following was concluded:

To maximize sulphur capture efficiency, the limestone grain size should be in the range of 40 to 600 microns.